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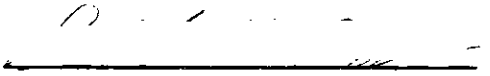

SOME CHEMICAL ASPECTS OF RAPID SAND FILTRATION

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SUMMARY

Conventional rapid sand filter design criteria deal almost exclusively with physical and mechanical filtration variables such as sand size, bed depth, and flow rate. The purpose of this study is to evaluate the chemical aspects of the rapid sand filtration process.

Four model filters were designed to ascertain the effects of various selected chemical systems on the removal of hydrous ferric floc. These effects were characterized by the rate of filter bed clogging as measured by head loss increase and rate of penetration of floc into the filter bed.

The effects of 50 mg/l of NaCl, Na_2SO_4, Na_3PO_4, and $MgCl_2$ in demineralized water were evaluated. Later studies employed aqueous systems containing 25 mg/l of chloride, sulfate, and phosphate ions at pH levels of 5.0, 7.0, and 9.5. Effects of these latter systems on the filtration process were compared with filters receiving ferric floc suspended in demineralized water.

Microelectrophoretic mobility measurements were employed to determine the effects of these various chemical systems on the surface characteristics of the ferric floc particles and floated silica sand. Improved filtration was correlated with decreasing negative mobilities of the flocculant suspensions. Poorest filtration was observed at pH 9.5 for all systems analyzed. Phosphate ions were found to have very significant effects on filtration at all pH levels.

A thorough review of possible particle collection mechanisms is

presented. These mechanisms are discussed in connection with the results of these and other filtration studies.

This study has resulted in the following conclusions:

1. The removal of ferric floc from suspension during rapid sand filtration as evaluated by head loss increase and rate of bed penetration can be significantly affected and, in fact, controlled by the chemical composition of the aquecus suspending medium.
2. Electrokinetic effects resulting from the charged surfaces of the sand filter medium and the suspended floc particles as determined by microelectrophoretic mobility measurements provide the most important particle collection mechanism in the rapid sand filtration of hydrous ferric oxide floc.
3. Phosphate ions exert very significant effects in the filtration of ferric floc, as characterized by a marked decrease in clogging rate and a corresponding increase in the rate of floc penetration into the filter bed.

CHAPTER I

INTRODUCTION

Water filtration may be described as the process by which water is separated from suspended matter by passage through a porous medium, usually sand. Although this process has been used in water purification for some 4,000 years[1], particle removal mechanisms during filtration have not been clearly defined. This research was undertaken in order to provide a better understanding of the filtration process and to improve sand filter design.

The "rapid sand filter" has gained the widest acceptance in this country for the filtration of municipal and industrial water supplies. Rapid sand filtration may be defined as a process in which water is passed through a porous bed at a rate of at least 2 gpm per sq ft of filter surface. Filter beds up to three feet in thickness are employed, and cleaning is accomplished hydraulically.

Normally, raw water receives some sort of pretreatment prior to filtration. One or more of a number of processes may be employed depending upon the characteristics of the raw water and the quality desired after treatment. In many cases some type of coagulation process followed by sedimentation precedes filtration. The filter is then called upon to remove those materials remaining in suspension after the settling process. This suspended matter generally consists of a small fraction of the raw water turbidity together with an appreciable amount

of coagulant. In most cases iron or aluminum salts are utilized as coagulating agents and the filter is then called upon to remove appreciable quantities of the hydrous oxides of these metals. The average diameter of these floc particles applied to the filters has been found to be in the order of about 20μ and is considerably less than the average pore diameter in the filter bed.

It is postulated that physical removal mechanisms cannot adequately explain the removal of these materials from suspension during the filtration process. It is believed that chemical aspects of the rapid sand filtration process play a much more prominent role than indicated by previous investigators. This research and thesis are directed toward a test of this concept.

CHAPTER II

LITERATURE REVIEW

General

The earliest recorded instance of the filtration of impure water through sand is found in Susruta Sanhita, a medical treatise of the Hindus in India[1]. In a more recent era, slow sand filters were introduced in England in 1829, and rapid sand filters came into existence towards the end of the nineteenth century. Slow sand filters were introduced into the United States in 1872 with the first plant being built at Poughkeepsie, New York[2]. Municipal rapid sand filters followed with the construction of a plant at Somerville, New Jersey in 1885.

After the introduction of sand filters into this country, a long period followed during which attention was focused on refining the filter apparatus. A period of refinement in filter operation followed, after which emphasis was directed toward proper conditioning of the water prior to the filtration step. Only in recent years have significant efforts been made toward determination of the mechanisms of suspended particle removal during filtration.

The filtration process has been evaluated by previous investigators in terms of one or more of the following criteria: bed penetration, turbidity removal, head loss, and length of filter run. The importance of these criteria can be illustrated by considering a typical filtration run.

When turbid water is applied to a filter, the amount of turbidity diminishes as the water passes through the filter, resulting in a more or less clear filtrate[3]. As the filter run proceeds, the bed becomes clogged, and in many cases the filtrate at first improves. Later it steadily deteriorates, and ultimately, if sufficient pressure can be maintained, the filtrate becomes as turbid as the influent. The clogging of the filter is worst near the surface and as the run proceeds, gradually works down through the bed. The head loss depends on the degree of clogging and most of the pressure drop occurs in the upper layers of the filter. However, as the clogging extends downward, the lower layers also contribute to the head loss[3].

The primary purpose of a filter is to produce an aesthetically acceptable water. Since the operational limits of the filter are established by the design, it is very important that all parameters of filtration are considered during the design process.

The filter must remove turbidity to an acceptable limit at an economic rate. If all the turbidity is removed at the surface, the head loss is great, the length of filter run short, and the operational costs high. On the other hand, if the turbidity readily penetrates the bed, a low head loss results, but the filtered water is not acceptable and again the length of filter run is short resulting in high cost. The ideal filter would share the removal burden throughout its depth, produce water of acceptable quality, and operate for a reasonable length of time resulting in a reasonable loss of head.

Filter runs are usually terminated when one of the following occur: (1) all the allowable head loss is utilized; (2) the turbidity has pene-

trated the bed resulting in unacceptable quality of water. Often an arbitrary time limit is set for termination of filter runs since most filtration plants do not have provisions for measurement of effluent turbidity from each filter. Such a time limit may also be set to prevent penetration of bacteria when significant biological activity occurs within the filter bed.

Filtration Variables

Research by previous investigators has established that the following factors[4] can affect sand filtration:

1. bed characteristics
    a. sand size
    b. sand shape
    c. surface properties of the sand grains
    d. packing arrangement of the sand grains
    e. bed depth
    f. bed porosity
2. suspension characteristics
    a. particle size
    b. particle shape
    c. particle density
    d. surface properties of the particles
    e. chemical composition of the particles
    f. chemical composition of the water
    g. water temperature

3. operational characteristics

   a. flow rate

   b. backwashing procedure

A review of these factors reveals that many are physical in nature and are controlled by the design of the filter. Each is important but little can be done to affect filtration after the filter is constructed. The writer feels that these physical and operational variables have been thoroughly investigated and they will not be discussed here. O'Melia[4] has presented a thorough review of these parameters. The interested reader is also referred to the invaluable experiences of Baylis[5-15] and Hudson[16-21]. Chemical variables will be discussed in some detail later in the thesis.

Formulation of Filtration

According to Ghosh[22], the first progress made in characterizing the flow of water through granular beds was by Darcy in 1830. His studies on the rate of flow of water through beds of sand of various thicknesses resulted in the familiar relation:

$$V = K \frac{\Delta P}{L}$$

where ΔP = the pressure drop (ft) through the bed of thickness L (ft),

V = the average velocity of flow of the fluid (ft/sec),

K = a constant called the permeability which depends on the physical properties of the bed and the fluid (ft/sec).

This formula holds for all fluids flowing through granular beds in the laminar flow regime.

Considerable effort has been directed toward evaluation of K, the permeability coefficient. In 1892 Hazen[23] suggested the following empirical equation for approximating the permeability coefficient:

$$K = Cd_{10}^{2}$$

where d_{10} = Hazen's "Effective Size" of sand, which is the sieve size (mm) through which ten per cent of the material passes and 90 per cent is retained, and

C = a coefficient approximately equal to 100 but varied between 41 and 146 for sand sizes between 0.1 and 3 mm and a uniformity coefficient less than 5.

In 1927 Kozeny[24] extended the Poiseuille equation for laminar flow through capillary tubes and developed the following equation:

$$K = \frac{g}{K'\mu S^{2}} \frac{f^{3}}{(1-f)^{2}}$$

where K = permeability coefficient,

g = acceleration due to gravity,

S = specific surface of particles,

f = porosity of the media,

K' = a constant, equal to 5.0 for spherical particles, and

μ = viscosity.

Fair and Hatch[25] independently developed an equation of the Kozeny form in 1933. Carman[26], Coulson[27], and Rose[28,29] have also made significant contributions in the evaluation of flow of fluids through porous media.

The formulae presented so far correlate well for the laminar flow of clean water through randomly packed porous media. In filtration, however, turbid water applied to the filter results in continuous changes in the characteristics of the filter bed. These relationships, therefore, offer only a starting point.

Attempts to characterize the filtration process have resulted in numerous empirical formulae. Two of the most frequently qucted were formulated by the American Society of Civil Engineers[30] and by Stanley[31,32].

The ASCE formulation is for a terminal head loss of 8 feet, an effluent containing 0.2 mg/l of turbidity or less, and a constant rate of filtration of 2 gpm per sq ft:

$$L = Kd^{1.67}\left(\frac{60}{T + 10}\right)$$

where L = required filter depth (inches),

d = sand grain diameter (mm),

T = water temperature (degrees Fahrenheit), and

K = coefficient varying according to the water quality.

Stanley's formulation is for an 8 foot terminal head loss at 25 degrees centigrade:

$$P = Kd^{2.46}Q^{1.56}$$

where P = depth of penetration of ferric floc (inches),

d = sand grain diameter (mm),

Q = rate of filtration (gpm/sq ft), and

K = 6.4 for the units given.

Neither of these formulae give any indication of how either the quality of filtrate or the head loss will vary with time, or what will be the effect of changing the depth of the filter. Since they are empirical in nature, they can be successfully applied only under the same conditions from which they were derived.

Probably the first attempts at a rational analysis of the filtration process to correlate the time variation of the amount of turbid material remaining in suspension at a given depth were made by Iwasaki[33] in Japan in 1937. Basically, he proposed that removal of suspended solids through the depth of a filter is exponential. The following formulation mathematically describes this relationship:

$$C = C_o e^{-\lambda L}$$

where C_o = initial (surface) concentration,

C = concentration remaining at depth L, and

λ = filter coefficient.

The filter coefficient λ is dependent upon the nature of the suspension, the rate of filtration, water viscosity, and the internal geometry of the porous filter. Since the internal geometry changes during filtration, λ is not constant. Using the same basic equation, Mackrle

and Mackrle[34], and Ives[35] have attempted to evaluate filtration. Most of their efforts were directed toward evaluation of the filter coefficient, λ.

Mints[36] developed a formulation for the filtration process based on deposition on the filter grains on existing deposits, and breakaway of existing deposits by the flowing water. The differential equation developed states that the change (reduction) in concentration through a unit depth is proportional to the concentration at that depth less a proportion of the existing deposit returned to the flow by hydrodynamic shear.

Possible Particle Collection Mechanisms

A rational analysis of filtration requires that the primary particle removal mechanism(s) of the process be defined. Over the years many such mechanisms have been proposed but no agreement has been reached on this topic. The following discussion presents a summary of proposed mechanisms of particle collection within a filter.

The writer chooses to discuss particle collection mechanisms rather than removal mechanisms since the former can be better defined. For purposes of discussion consider particles in suspension of radius R. At any time that a suspended particle comes within a distance R of a sand surface, it shall be considered removed. Whether or not a particle remains or later penetrates deeper into the filter bed is itself a matter of controversy. Stanley[31,32] reports that the particle does not move deeper into the bed. On the other hand, Stein[37] and Mints[36] report that it does. In either case it shall suffice to say that a particle is indeed removed

when it comes within a distance R of the sand surface provided the "sticking forces" equal or exceed the hydrodynamic shear forces and any repulsive forces that may be present.

The following particle collection mechanisms have been proposed:

1. direct sieving or straining,
2. sedimentation,
3. inertial impingement and centrifugal collection,
4. Brownian movement,
5. van der Waals forces,
6. electrokinetic forces,
7. convergence of streamlines, and
8. diffusion due to concentration gradients of suspended particles.

1. Direct Sieving or Straining

One of the first mechanisms proposed for particle collection is straining or direct sieving. This theory postulates that a sieving action occurs in the crevices adjacent to points of contact between the sand grains[38]. In each of these angular areas normal to the flow there is an opening with a width smaller than the dimensions of any given suspended particle. The portion of flow, however small, that passes through this area will be stripped of all particles of that size or larger. Since flow is usually in the laminar range in sand filtration, little flow (if any) would be expected in these very small angular spaces. For this reason some writers consider the pores to form channels with varying radii through the filter bed. Straining is the undisputed collection mechanism for suspended particles larger than the constrictions within these channels. However, filters have been found to remove particles much smaller

than the average pore diameter within the filter bed. Investigators have postulated additional phenomena in which either the suspended particles become larger or the pores become smaller.

Coagulation within the filter bed could produce an increase in suspended particle size. In water treatment the probability of contact between particles is increased by gently mixing the suspension. Supporters of the coagulation theory in filtration picture the pores in the filter bed as small coagulation basins. Optimum filtration would then be expected to occur under conditions favorable for coagulation. Stanley[31,32] reported optimum filtration near pH 7 which he believed to be the isoelectric point of the ferric floc under study. He concluded that the best coagulating floc would probably also be the best filtering floc.

The rate of flocculation is a function of the mean velocity gradient[39]. Stein[37] reported that a velocity gradient approximately 40 ft/sec/ft is optimum for flocculation. Stanley[32] reported a velocity gradient of 38.5 ft/sec/ft for a filter with 20-30 mesh sand, 2 gpm per sq ft filtration rate, pore diameter of 0.5 mm and a porosity of 36%. He ruled out flocculation (even though having a favorable velocity gradient) on basis of detention time since the total amount of flocculation is a function of the mean velocity gradient times the time[39]. If most material were removed in the top inch of filter a detention time of only 20 seconds results which he contended is too short.

Conley and Pitman[40,41] have reported on the use of coagulant aids as filter aids. They report better removal of turbidity by the filter after adding coagulant aids. They have attributed this action to better "sticking" of the turbidity to itself and the sand grains.

Garnell[42] has reported increased removal of turbidity by a filter by adding a polyelectrolyte just prior to filtration. Addition of 10-30 ppb of polyacrylamide results in a greater head loss increase and shorter runs but increased removal efficiencies. Black[43] has discussed the role of such polyelectrolytes in coagulation.

The coagulation process is discussed at length in the section on electrokinetic forces.

Investigators of the filtration process have noted the development of a visible layer of filtered material on the surface of the filter medium. This layer, termed the Schmutzdecke (dirty layer), was found to contribute greatly to the head loss across the filter. Various investigators have attributed different degrees of filtration to this layer. Hardenbergh[44] reported straining due to the Schmutzdecke to be the major removal mechanism. Stein[37] did not believe it to be of any real importance to filtration. He based this conclusion on studies of the filtration process with the use of a "microfilter." This small filter was equipped with a microscope and the filtration process could be visually followed. He concluded that removal by the Schmutzdecke is negligible since high velocities result when the small pores at the surface become clogged. The result is penetration into the filter bed.

Some early investigators of the filtration process attributed removal of very small particles to the development of a "colloidal mesh structure" within the filter medium which effectively reduced the size of the pore openings in the filter bed. This mesh structure, it was believed, was primarily composed of organic matter of a gelatinous character forming a porous framework within the interstices of the filter.

It was noted that a sand filter is more effective in removing suspended material after it has been in service for a period. This "ripening" period, some investigators believed, was the time required for development of the colloidal mesh structure. With waters of high organic content the ripening period was shorter than that required for waters with little or no organic matter. It was believed that bacterial action[39,45] was important to the development of a colloidal mesh structure for slow sand filters. Rapid sand filters were frequently backwashed thus disturbing the mesh structure. Chemical precipitation, aided perhaps by the electrical charges[45] attributed to the colloids and sand grains, was believed to form the mesh structure in the rapid sand filter.

2. Sedimentation

The sedimentation theory pictures the small voids as miniature settling basins in which gravitational settling occurs as a result of the differences in densities of suspended particles and water. If this theory is valid, better filtration would be expected with an increase in particle density. Camp[46] has presented a thorough study of the sedimentation process and the design of settling basins. He has shown that theoretically the efficiency of sedimentation is independent of the depth and detention time. The ideal basin would have a depth approaching zero. While this is not practical a filter would come very close to approaching this ideal.

Hazen[47] has shown that a sand filter composed of uniform sand 0.5 mm in diameter would be expected to remove, by settling, particles 1/20 of the diameter and 1/400 of the settling velocity of particles removed in sedimentation basins having equal hydraulic loading. He arrived at this

figure assuming only 1/18 of the total sand grain surface to be available for sedimentation.

Hall[38] developed a differential equation for filtration based on the sedimentation theory. He reported good correlation with data presented by the ASCE Committee on Filtering Materials[48].

Stein[37], in his studies of the filtration process with a microfilter, concluded that sedimentation of suspended floc particles within a filter bed is not an important mechanism of collection. He observed best removal at points of maximum velocity which does not concur with the sedimentation theory.

Cleasby and Baumann[49] concluded that straining and sedimentation are the predominant removal mechanisms for calcium carbonate crystals. This conclusion was based in part upon microscopic examination of the sand beds after filtration. These crystals were found to predominate on the upper surfaces of the sand grains thus pointing toward sedimentation and straining as predominant collection mechanisms.

Ives[50] has developed a theory of filter behavior based on the following three assumptions:

1. the removal of suspended material from water by a filter layer is proportional to the quantity of suspended material present in the water,
2. the removal characteristics of the filter depend on the surface available on the filter grains, on the tortuosity of flow within the pores, and on the interstitial velocity, and
3. the principal force operating to remove suspended particles from the flow stream lines is gravitational, although adsorptive forces become

dominant when a particle reaches a given surface.

Ives developed an equation of filtration for which numerical solutions have been obtained on automatic digital computers. In order to test the validity of his equations Ives filtered suspensions of radioactive algae of the genera Chlorella and Scenedesmus. The distributions of the algae deposited in the filter pores were measured with a scintillation counter. He reported good correlation between his theory and test results. He offered, in support of his theory, that more Chlorella are removed per centimeter of depth of filter than Scenedesmus which is in accord with the relative settling rates of these algae in quiescent water.

In a later publication[51] Ives notes that this gravitational hypothesis cannot explain the high degree of removal normally noted for coagulant materials with an effective specific gravity close to that of the suspending medium.

3. Inertial Impingement and Centrifugal Collection

It has been suggested that the tortuosity of flow within the pores of the sand filter can contribute to inertial impingement of the suspended particles on the filter medium. If the mass of the particles and the velocity of flow are sufficiently high, the paths of the centers of the particles will not coincide with the stream flow lines of the fluid[52]. The flow lines have to pass around the sand grains and are therefore highly curved near the surface of a sand grain. However, the paths of heavy particles which are headed towards the sand surface do not curve away so rapidly and therefore the particle may come in contact with the sand grain even if the flow line which it started to follow does not come within the

distance R from the surface.

Langmuir[52], Davies[53], and Thomas[54] have investigated this collection mechanism for air filters and present relationships for its evaluation. Decker et al.[55] report the removal of suspended materials during air filtration is presently accomplished largely by inertial impingement.

Grace[56] used the relationships developed by Langmuir[52], Davies[53], and Thomas[54] to compare the relative magnitude of several of the commonly proposed collection mechanisms for the filtration of ferric oxide particles in a glycerol-water suspension by fibrous filter media. He reported the effects due to inertial impingement to be neglible when compared to other mechanisms.

Stanley[31] states that in rapid sand filtration the forces due to inertia are in the order of 1% of the magnitude of those due to gravity. He reasoned that if inertial forces are the contributing factor to better filtration, an increase in filtration rate would decrease the penetration of material into the bed. Since the converse is true, he concludes that inertia is not of importance as a removal mechanism.

Closely associated with inertial impingement is centrifugal collection[51]. As the fluid flows around a sand grain (curvilinear flow) a centrifugal force results which causes the particle to diverge from the flow line. If this divergence brings the particle within a distance R of a sand surface the particle is removed. This theory has been effectively applied in design of equipment for removal of suspended particles from air[53]. Increasing particle density and flow velocity would result in better filtration if this mechanism were important.

4. Brownian Movement

Brownian movement is that motion imparted to the suspended particles by randomly moving ions and molecules in the solution. Brownian movement would be of importance only for small particles, but since theories have been sought to explain the removal of such particles it has been suggested as a collection mechanism. The centers of suspended particles effected by Brownian movement do not move exactly along the flow lines but wander irregularly from positions on the flow lines[52]. This diversion from the stream line increases the probability of the particle coming within the distance R of the sand surface.

Stanley[31] has calculated it would take 20.3 seconds for a particle only 10^{-5} cm in diameter to move 10^{-3} cm due to Brownian movement. During this time roughly one-half of the particles of this size would be removed. He disregarded Brownian movement as a removal mechanism since he reasoned that a very small percentage of particles would be within 10^{-3} cm of a sand surface. Grace[56] reported the effects due to Brownian movement are negligible when compared to the other proposed removal mechanisms.

5. van der Waals Forces

The van der Waals forces are forces of attraction that may be described as molecular cohesive forces that increase in intensity as the particles approach each other[43]. These forces between two atoms will be proportional to r^{-7}, which means a rather rapid decay with increasing distance between the atoms[57]. For large particles separated by short distances the forces are proportional to r^{-3}.

Mackrle and Mackrle[34] developed a dimensionless group, known as

the "adhesion criterion", based on the van der Waals and hydrodynamic forces acting within a filter. They reported that with a knowledge of the hydrodynamic characteristics of the filter combined with certain empirical data on the floc being used, the reduction in suspended solids can be calculated for filters of any media grading and of any given depth.

6. Electrokinetic Forces

Electrokinetic forces have been studied most in connection with coagulation. A better understanding of these forces is gained by a review of this process.

The principal function of coagulation[43] is the destabilization, aggregation, and binding together of colloids. The two most important instability factors for colloidal systems are the Brownian movement and van der Waals forces of attraction. These two factors are opposed by two stabilizing factors: the zeta potential and degree of hydration of the colloids.

The translation due to Brownian movement increases the probability of a collision between two colloidal particles. The van der Waals forces of attraction may be responsible for coagulation if the distance between the colloidal particles is sufficiently reduced by Brownian movement or other means.

Hydration is a property possessed by some particles enabling them to attract relatively large numbers of solvent molecules to their surface[43]. Water molecules are polar, and hydrophilic (water loving) colloids attract a layer of these molecules to their surfaces. This hydrated layer hinders contact between the colloidal particles.

All solid particles have a charge at their surface when placed in contact with water. This charge can be due to one or more of the following:

(1) ionization of molecules at the particle surface,

(2) unsatisfied charges due to imperfections in the crystal lattice,

(3) direct chemical reaction (chemisorption) with specific ions in the solution with the formation of chemical bonds, and

(4) weaker, physical adsorption of ions from solution as due to hydrogen bonding and van der Waals forces.

The presence of this surface charge attracts ions of opposite charge from the solution. There is produced a very tightly held layer of ions of opposite charge called the "stationary layer" and a second more loosely held layer (also of opposite charge) called the "diffuse layer." The two layers together result in zero net potential for the particle. This diffuse-double-layer theory[43] attempts to explain the behavior of colloids.

If an electric field is applied to the system, the particles will migrate toward the pole of opposite sign. The stationary layer migrates with the particle and the diffuse layer is "sheared" off forming a plane of slip at the stationary layer. The potential difference between the stationary layer and a point of zero potential is the zeta potential.

If an indifferent electrolyte is added to a colloidal suspension, the zeta potential is effectively reduced. As the concentration of electrolyte is increased more ions enter the stationary layer leaving less charge to be neutralized by the diffuse layer. For this reason electro-

lytes are said to contract the diffuse layer. If a sufficient amount of electrolyte is added, a point will be reached when the van der Waals forces of attraction will overcome the repulsive forces of the zeta potential and rapid coagulation results[43].

The valence of the ion (in the electrolyte) of opposite charge to the colloid is very important. Verwey and Overbeck's theory[57] indicates that a bivalent ion is 64 times more effective than a monovalent ion in coagulation while the trivalent ion is 729 times more effective. Coagulation can also be affected by mixing two colloids of opposite sign[43]. This is known as mutual coagulation.

This summary of the coagulation process is somewhat simplified but will suffice for the purposes of this thesis. Black[43] and Packham[58-60] have presented excellent discussions on the theory of the coagulation process. Stumm and Morgan[61] have presented a thorough review of the chemical aspects of coagulation. Weiser[62-64], Hazel[65,66] and their respective co-workers have contributed much to the knowledge and understanding of the effects of electrolytes on the coagulation process and mutual coagulation. More recently, Black and co-workers have presented a series of outstanding papers[67-73] describing microelectrophoretic investigations of the coagulation process. The works of Verwey and Overbeck[57] and van Olphen[74] were found to be most helpful in gaining an understanding of coagulation and the role of electrokinetic forces.

In filtration these electrokinetic forces will be present on the surfaces of both the suspended particles and the sand grains. Ellms[45] in 1928 described the filtration of ferric hydroxide suspensions, and noted that the floc was retained for a period of time after which it passed

through unchanged. It was assumed that the positively charged colloidal particles of ferric hydroxide were retained by the negatively charged sand grains.

Stanley[31] reported that the negatively charged grains are coated very rapidly by positively charged particles resulting in a charge essentially the same as that of the suspended particles. This would tend to negate the importance of the sand surface charge. A floc very close to the isoelectric point would allow optimim coagulation of suspended floc particles which he believed to be of more importance than the surface charge of the sand.

Sanford and Gates[75] conducted experiments employing a normal sand filter which had been chemically treated with a stearato chromic chloride to produce a hydrophobic stearate coating on the sand. This coating was assumed to have produced a positively coated sand surface. Their results showed no differences in the removal of either hydrous aluminum oxide floc (assumed to be positively charged) or Escherichia coli (negatively charged) by the two filters. It was concluded that adsorption is not of importance in the removal of these materials.

Cleasby and Baumann[49] investigated the filtering properties of waters containing hydrous ferric oxide floc. They concluded that electrokinetic forces were primarily responsible for particle removal. This conclusion was based in part upon microscopic examination of the sand beds after filtration. The ferric floc particles exhibited no preference for either horizontal surfaces or pore interstices, thereby implying adsorption onto the sand surfaces.

O'Melia[4,76] investigated the filtration of algae of the genera

Scenedesmus, Anabaena, and Ankistrodesmus. These algae (which are negatively charged) passed readily through the model filters. Addition of iron coagulant greatly improved the algal removal efficiences. Electrokinetic forces[76,77] appeared to be of primary importance to the removal of these materials from suspension.

Ives[3] also studied the filtration of algae. He reported continuous passage of algae through the filters.

Viruses, algae, and bacteria (negatively charged) are known to be removed more efficiently if coagulants[4,78,79,80] are used. This points toward the possible importance of electrokinetic forces.

Oulman and Baumann[81] studied the effects of zeta potential of the filter medium in diatomite filtration. They determined the effects that several inorganic ions have upon the zeta potential. In a later publication[82] a formulation for diatomite filtration is presented in which the zeta potential is not taken into account. This indicates that they do not place much importance in this parameter.

Jordan[83] conducted a study of the filtration of clay suspensions at different electrophoretic mobilities, controlled by the addition of sodium hexametaphosphate. Gravel filters were operated for 65 days in order to develop a Schmutzdecke. Improved filtration was then noted with an increase in negative mobility of the clay particles. Jordan offered as a possible explanation the presence of a positively charged Schmutzdecke.

7. Convergence of Streamlines

As a suspension flows through a filter there are many constrictions through which it must flow. The streamlines converge at each constriction. Stein[37] observed a buildup of floc at the constrictions in his

microfilter apparatus. He reasoned that the removal was due to a "contact action" produced by the convergence of the fluid streamlines. An equation was developed for an idealized contact mechanism based on the convergence of streamlines at the constriction of a tube. It was concluded that the probability of removal of a particle in a unit depth varies directly as the square of the diameter of the suspended particles and inversely as the cube of the diameter of the sand grains.

8. Diffusion Due to Concentration Gradients of Suspended Particles

Particles are known to diffuse from areas of high to those of lower concentration. Hunter and Alexander[84] have proposed that this diffusion into "dead spaces" within the filter medium could explain removal of particles much smaller than the average pore diameter of the filter. Their results are based on studies of flow of kaolinite sols through silica sand columns. They contend that the shear rate is low in the dead spaces; nevertheless, the yield strength of the sol must be sufficiently strong to resist the shearing forces that are applied by the moving liquid. Otherwise, the particles would be removed and thus penetrate deeper into the filter.

They studied the effect of zeta potential of silica on deposition of kaolinite within the filter[84]. Streaming potential measurements on the silica column showed that the zeta potential of the silica surface was similar to that of the clay under the experimental conditions studied. When the silica surface was rendered positive by the adsorption of hexadecyltrimethylammonium ions, the general features of the deposition process were unchanged though the amount of deposition was substantially increased.

Directly opposed to the diffusion of particles due to a concentration gradient is the concentration of particles along the axis of flow[85-87]. There is a tendency for particles to migrate toward a region where the velocity gradient is a minimum[85] (therefore, toward the center of a tube) and the process is shear dependent. Eventually some sort of equilibrium would probably be reached between these two phenomena. More knowledge is needed to properly evaluate this mechanism.

Streaming Potential Effects

Head loss is an important consideration in filtration and was covered briefly in the general discussion of the process. This loss of pressure during filtration is due to clogging of the pores within the filter bed[88] and possibly to electroviscous drag due to establishment of a streaming potential[89,90] across the filter bed. Loss of head due to clogging of the pores is easily understood and readily acceptable but there is some controversy over the effects of streaming potentials.

Ruth[90] described head loss due to electrokinetic phenomenon as follows:

> When fluid starts to flow through a capillary, the velocity distribution profile near the wall is at first substantially parabolic in the direction of flow. Since no potential difference exists to cause migration of ions, fluid movement in the double electrical layer is without restraint. This results, however, in discharging a portion of the double layer from the end of the capillary, whereupon a small potential difference between the ends of the capillary immediately appears. Under the influence of this small e.m.f. positive ions adjacent to the wall at once begin to move toward the inlet end of the capillary. As the excess of positive ions in the effluent continues to increase, the potential becomes larger and the electro-osmotic flow greater. With an increasing electro-osmotic pressure opposing the mechanically applied pressure, fluid motion in the outer part of the double electrical layer adjoining the wall is first halted and then reversed in direction. The sine-wavelike profile characteristic of electro-osmotic flow against hydrostatic pressure increases

> in amplitude, at the same time the null point at which velocity again returns to zero moves out from the wall. Equilibrium is finally attained when sufficient potential difference is built up to transport ions electro-osmotically back to the inlet at the same rate they are swept from the inner portion of the electrical double layer by hydrostatic flow.

Ruth[90] demonstrated a decrease in head loss during cake filtration with an increase in sodium chloride concentration. He attributed this to a decrease in streaming potential in the filter with an increase in ionic strength of the filtrate.

Zaghloul[89] investigated the effects of streaming potential on filtration. In general he found the effects of streaming potential on head loss to increase with a decrease in sand size and an increase in flow rate. For sand sizes and flow rates commonly used in rapid sand filtration the effects were negligible. However, his investigations were made with clean water so that there was no clogging of the pores of the sand.

Oulman and Baumann[81] report that streaming potentials probably do not affect the build-up of head loss in a diatomite filter to any great extent.

CHAPTER III

CONSTRUCTION AND OPERATION OF MODEL FILTER APPARATUS

Construction

The experimental equipment employed in this research was designed to test the concept that chemical aspects of the filtration process are of prime importance in particle removal. It was desired to design and construct apparatus to provide controlled variation in the chemical composition of aqueous floc suspensions while maintaining physical and mechanical variables at constant levels. A detailed description of the model filter apparatus follows.

Figures 1 and 2 are photographs of the experimental apparatus used in this research. The apparatus was located in the hydraulics laboratory of the Civil Engineering Department at Georgia Tech. Figure 1 shows the portion of the apparatus installed on the second floor of the lab while Figure 2 shows that located below on the first floor. A schematic diagram of the entire apparatus is shown in Figure 3.

A demineralized water storage tank with a capacity of 500 gallons was constructed of plywood and provided with an inside coating of fiberglass to insure water tightness. The purpose of this storage tank was to minimize temperature variation and to provide an adequate supply of demineralized water for the duration of a run. A piezometer was placed on the outside of the tank so that the water level could be easily observed. City of Atlanta tap water was stored in the tank after being de-

Figure 1. Experimental Apparatus, Second Floor Level.

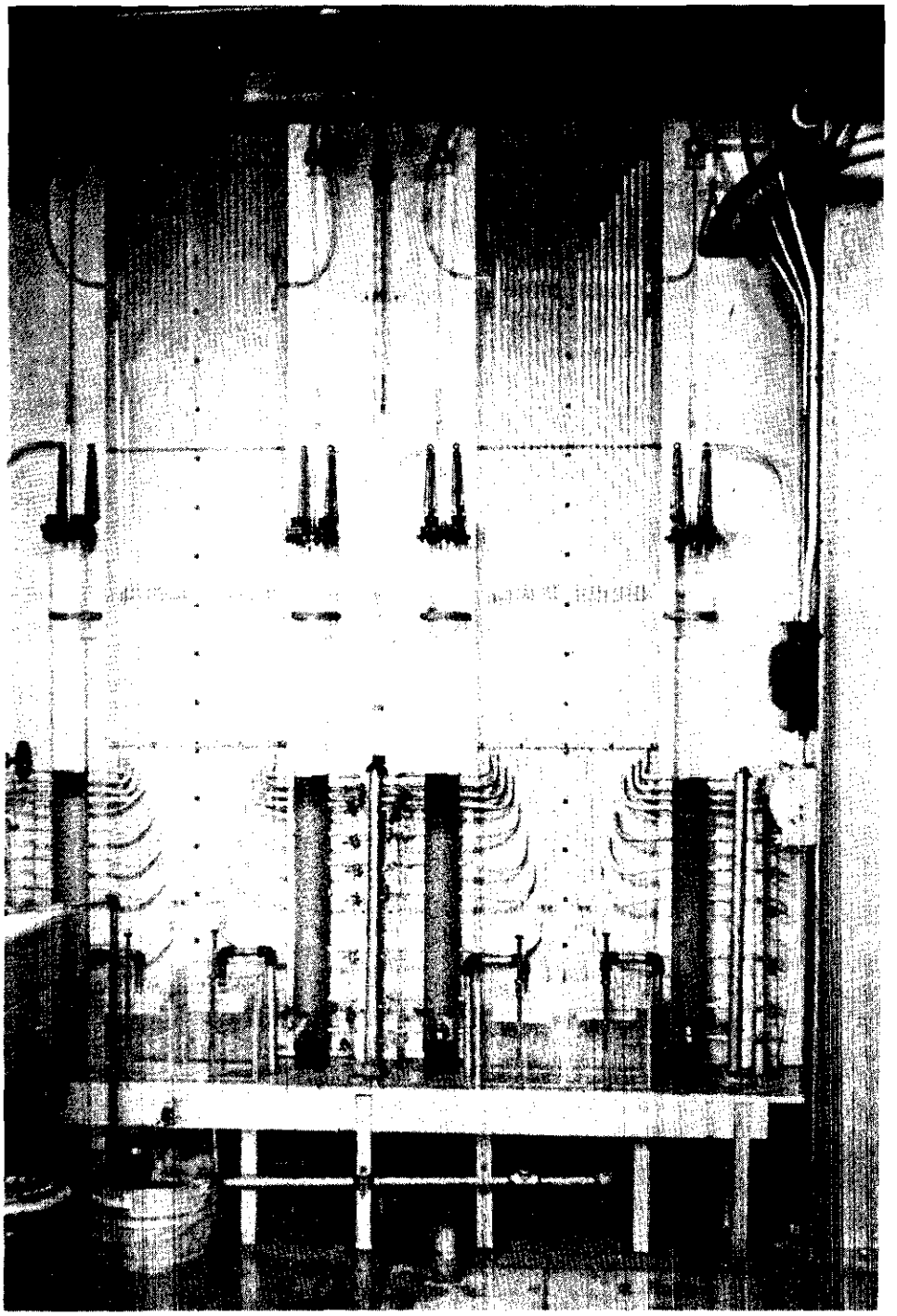


Figure 2. Experimental Apparatus, First Floor Level.

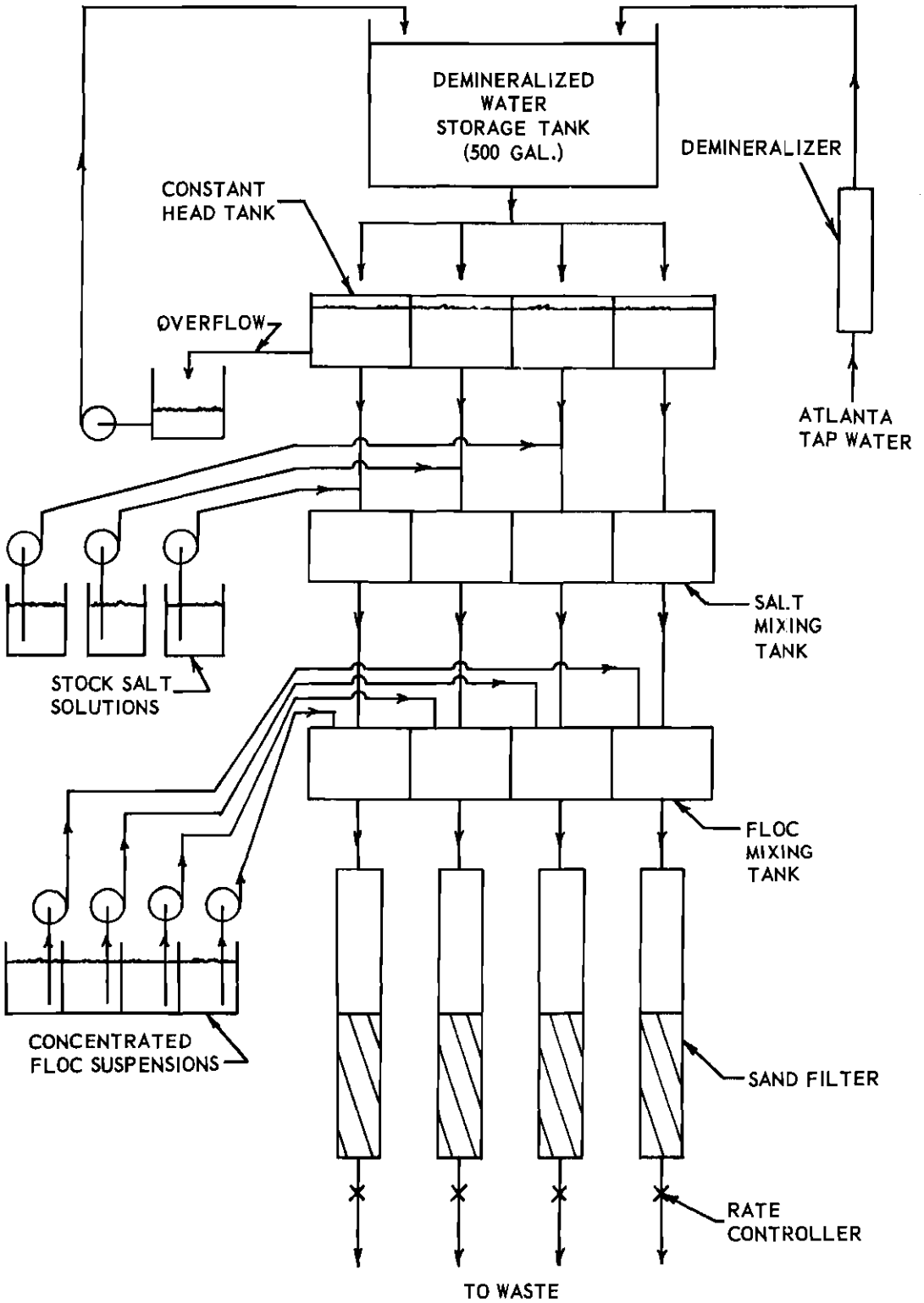


Figure 3. Schematic Diagram of Filter Apparatus.

mineralized by passage through a small mixed bed ion exchange unit installed in the supply line. The tank was filled at least one day prior to the start of a filter run to enable the water to reach room temperature.

Water was withdrawn from the storage tank through a 1/2 inch copper pipe, split into four streams by a manifold constructed of 3/8 inch copper tubing, and transmitted to the constant head tanks. A detail of these constant head tanks is shown in Figure 4. The flow rate from the storage tank was adjusted by a 1/2 inch gate valve in the copper line. Screw clamps were employed to regulate the flow split produced by the manifold. The system was regulated to provide an inflow to each constant head tank that was greater than the flow rate desired for filtration. Excess flow was diverted over the weirs in the constant head tanks, transmitted by gravity flow to a sump, and then returned to the storage tank by a small float operated sump pump. The constant head tank was constructed from 1/4 inch thick plexiglas.

The effluent from each constant head tank was transmitted through 3/8 inch white gum rubber tubing to the salt mixing tanks, presented in detail in Figure 5. Stock solutions containing selected inorganic salts were pumped from storage carboys into the salt mixing tanks through a glass "Tee" placed just upstream from each tank inlet. The tanks were filled to a six inch depth with 1/2 inch Raschig rings to provide thorough mixing of the stock salt solutions with the demineralized water. A Sigmamotor positive displacement pump with a variable speed drive was employed to pump the salt soltuions to the mixing tanks.

The effluents from the salt mixing tanks, representing four dif-

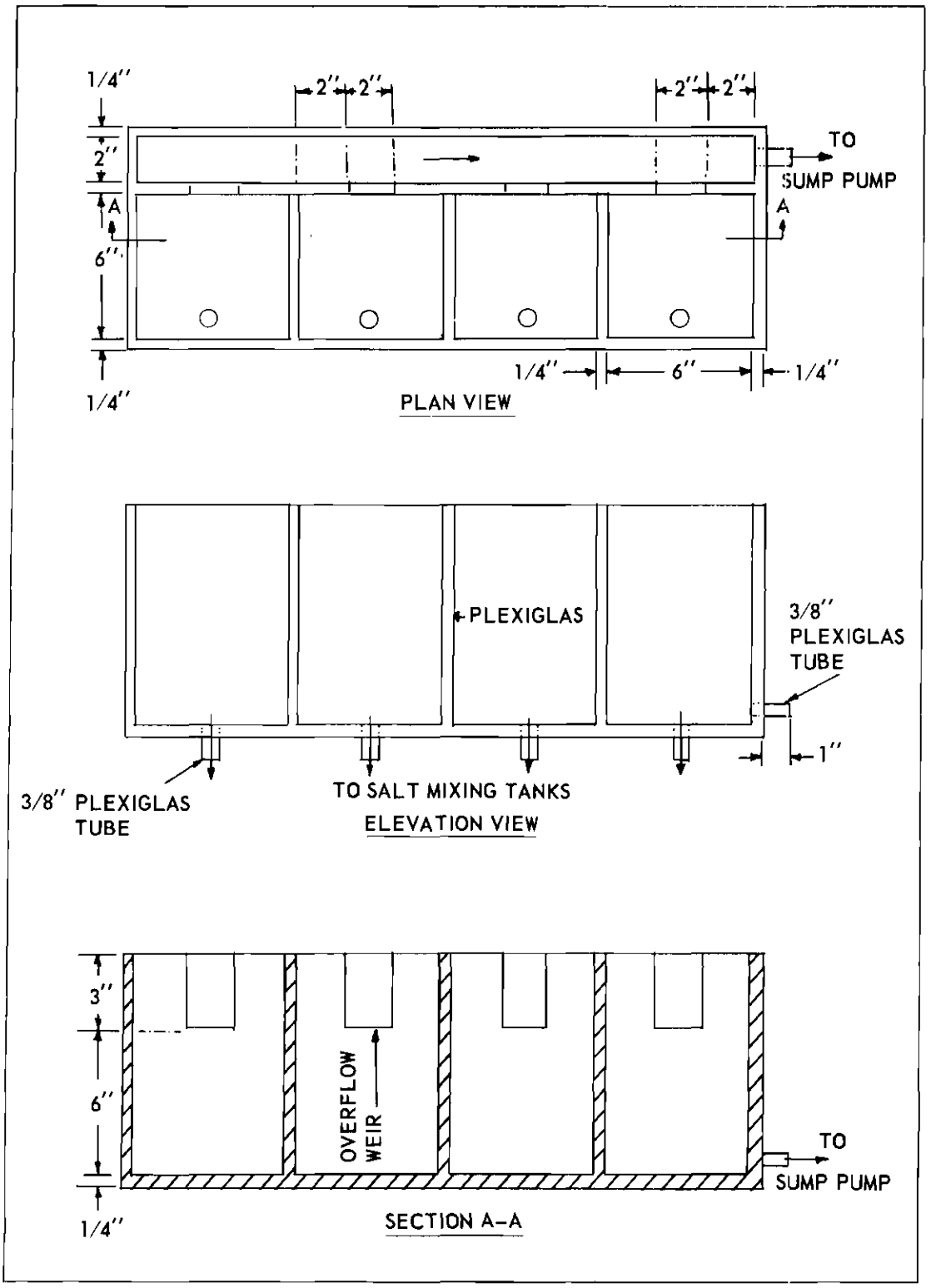


Figure 4. Detail of Constant Head Tanks.

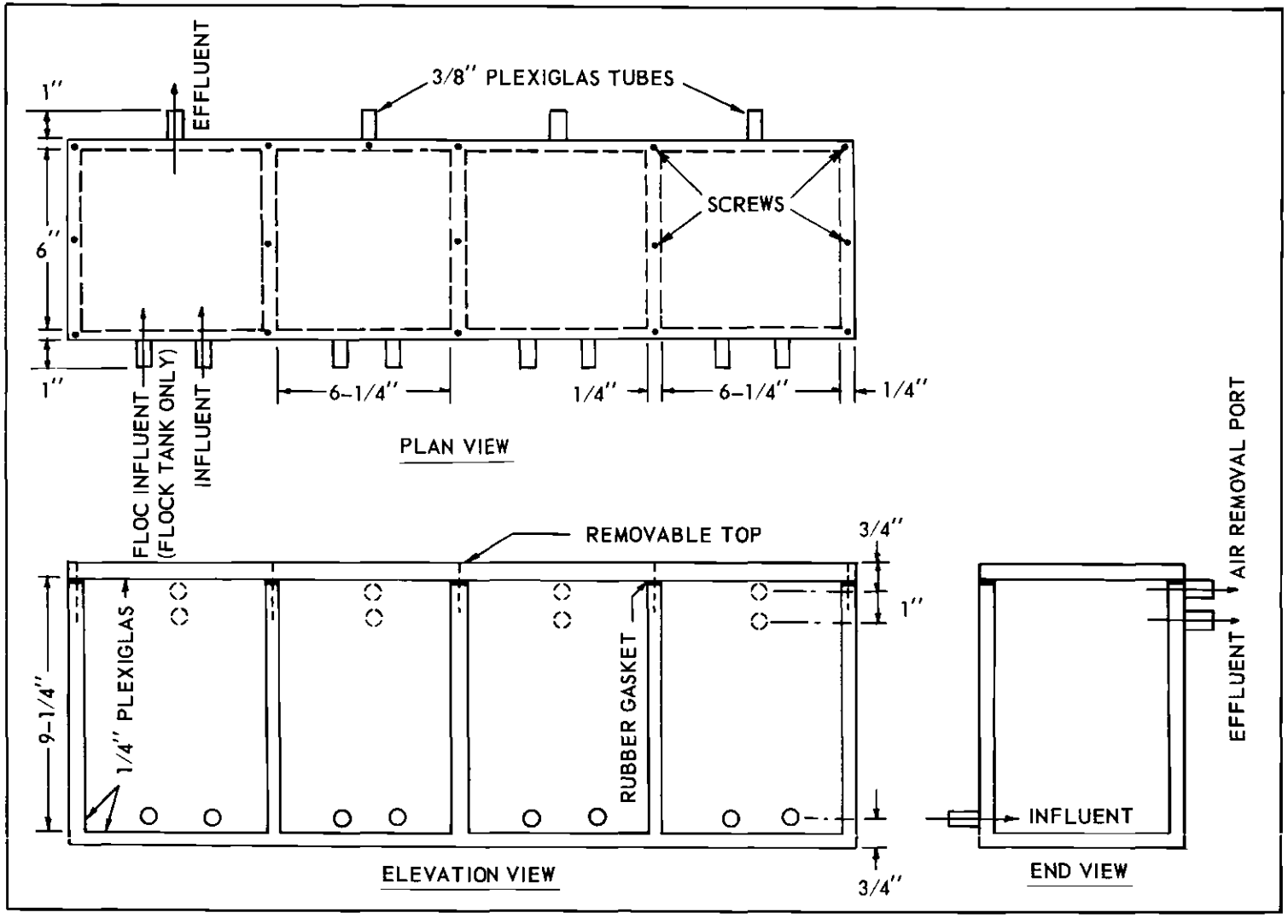


Figure 5. Detail of Salt and Floc Mixing Tanks.

ferent aqueous systems, were conveyed by gravity flow through 3/8 inch white gum rubber tubing to the floc mixing tanks. These mixing tanks were identical in size and shape to the salt mixing tanks, and their details can also be obtained from Figure 5. Mixing was provided by magnetic stirrers. Concentrated floc suspensions were pumped from storage tanks into the floc mixing tanks by four positive displacement pumps mounted in parallel on a single shaft and driven by a variable speed motor. Flow through each pump could be adjusted within the range of 0 to 30 milliliters per minute by adjusting the speed of the motor.

The concentrated flocculent suspensions were introduced into the mixing tank at the bottom, just above the magnetic stirring bars. The resulting floc mixture was withdrawn from the tank on the opposite side near the top. In addition to the floc and water inlets and the outlet, an additional outlet was provided near the top to facilitate the removal of air when the tanks were being filled. These compartments provided a theoretical detention time of about fifteen minutes when the filters were operating at a rate of 2 gpm per sq ft.

The stock concentrated floc suspensions were agitated throughout the course of a filter run in a specially constructed plexiglas storage tank with four 8 inch x 8 inch x 6 inch compartments. A variable speed stirring apparatus similar to the standard "jar test" equipment was constructed to provide the necessary agitation. The procedures employed in preparing these stock suspensions are presented in detail in Chapter V.

The effluents from the floc mixing tanks, representing four floc suspensions in different aqueous systems, were transmitted through 3/8 inch plexiglas tubing to four laboratory sand filters. The model filters

were constructed of plexiglas tubes having an inside diameter of 3 inches and a wall thickness of 1/4 inch. A four-foot section was flanged at each end to form the filter column. The 3/8 inch influent line was also flanged and then bolted to the top of the column. The bottom of the column was attached to an additional 5 inch length of the 3 inch diameter tubing with another flanged connection. A recess of approximately 1/16 inch was provided between the upper end of the 5 inch plexiglas tube and the face of its flange in order to accomodate the filter underdrain. Designed to provide support for the sand column, the underdrain was constructed by soldering 100 mesh screen to a copper ring having the same cross section as the plexiglas tube. Water tight connections were facilitated through the use of "O-rings." A large rubber stopper was fitted into the lower end of the filter column. A 1/2 inch pipe was provided to transmit filtered water from the filter column to the rate of flow controller. A detail of a model filter is presented in Figure 6.

The filters were designed to accommodate 24 inch deep sand beds. Manometer and sampling points were placed at the sand surface and at depths of 1, 2, 3, 6, 9, 12, 18, and 23 inches, and on the effluent side of the underdrain. Details of the manometer and sampling connections are shown in Figure 7. Sampling tubes were connected by 1/8 inch tygon tubing to a two inch length of 0.5 mm capillary tubing. A screw type tubing clamp was employed to insure flow control during sampling. Provision was made to prevent any distrubance of the sampling points and the filter bed during the sampling process. Manometer points were connected to 10 mm glass tubing attached to a central manometer board.

Rate of flow controllers were constructed utilizing a float op-

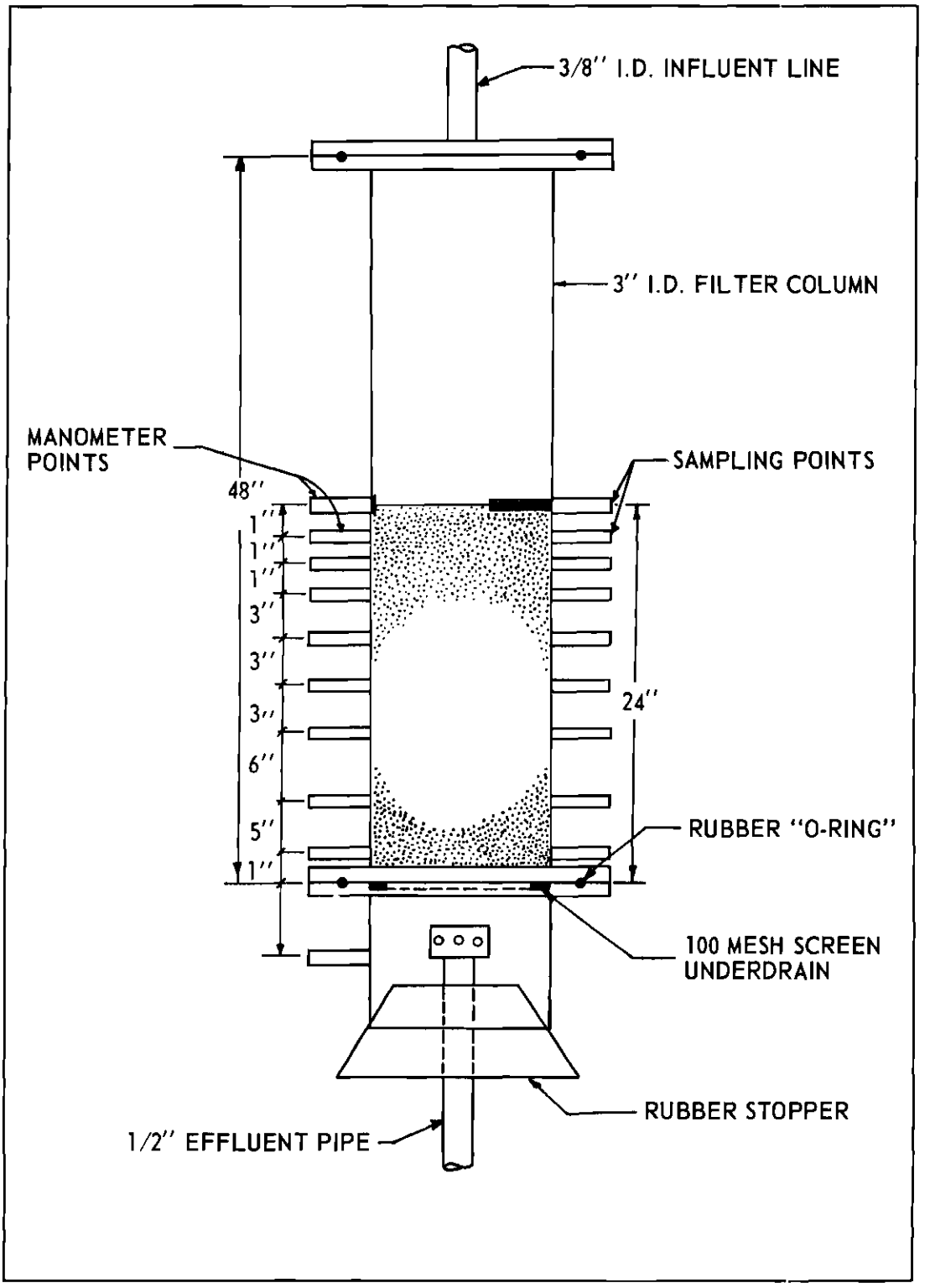


Figure 6. Model Filter.

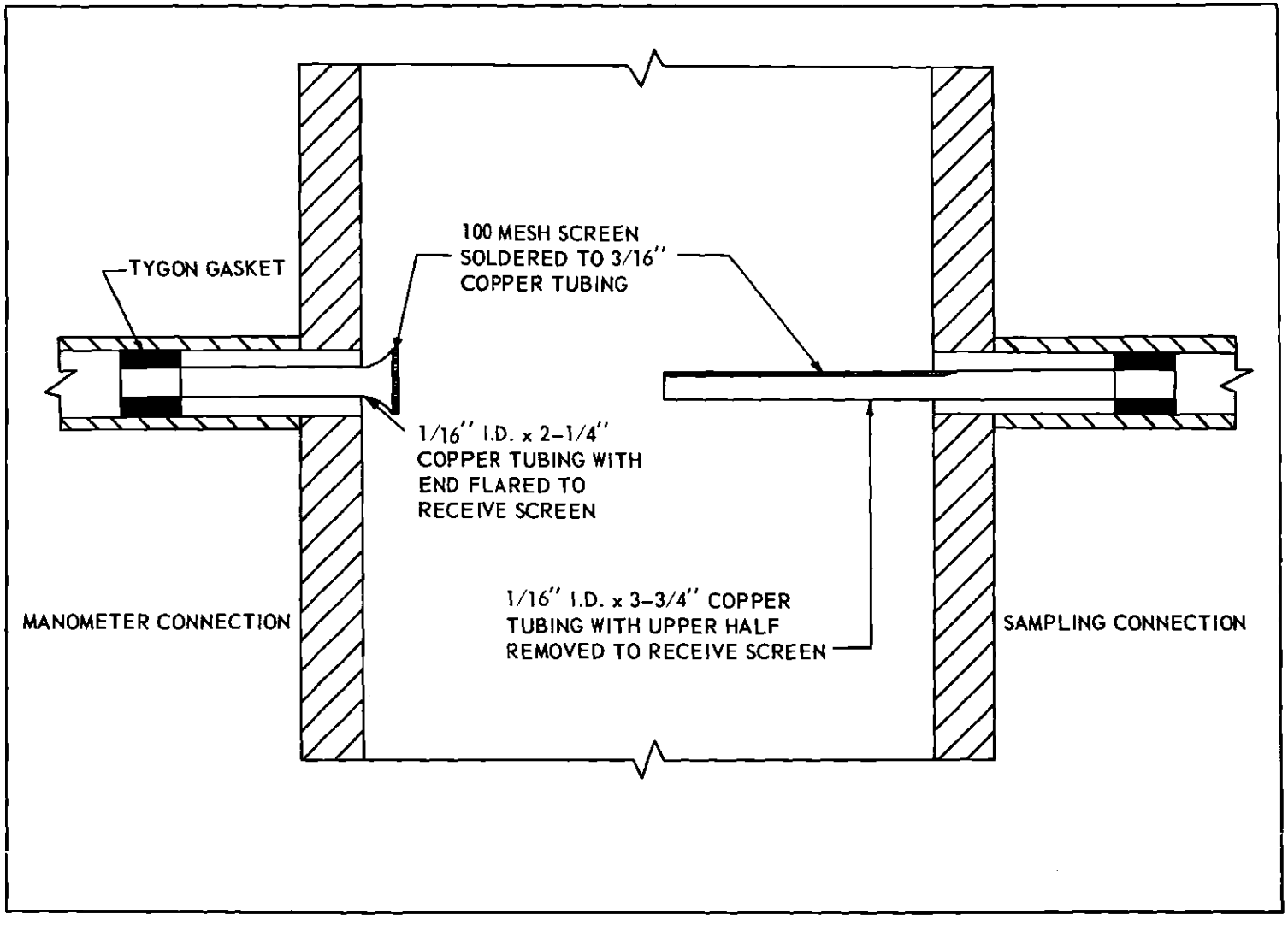


Figure 7. Detail of Manometer and Sampling Tube.

erated needle valve and a siphon. Details of a rate controller are shown in Figure 8. Each needle valve was constructed from a three inch length of 3/4 inch round brass stock and a 4-3/8 inch length of 1/4 inch round stock. The needle was supported on a paraffin coated styroform float. A 1/4 inch thick sheet of plexiglas was fastened to the top of the float to prevent its rotating in the float chamber. A thin copper sheet was attached to the bottom of the float to provide a firm mount for the needle. Originally, water was allowed to flow down the needle and over the surface of the float. It was noted, however, that a layer of water would build up on the float surface and then roll off, resulting in oscillations in the position of the needle and consequent sudden changes in flow rate. Provision was then made to allow the water to flow downward through the middle of the float by drilling holes through the plastic cover, the styrofoam, and the copper plate. The taper of the needle was adjusted to provide optimim flow control without sudden fluctuations due to slight movement of the float occasioned by jarring or vibration. After several trials, a ten degree taper was found to give excellent results.

An adjustable glass siphon was employed to deliver the desired flow from the float chamber. Siphons were constructed from 4 mm glass tubing, with the discharge end fire-polished to form an orifice. The head on the siphon (and consequently the flow rate from the float chamber) was adjusted by raising or lowering the siphon, a process which could be closely regulated by merely rotating a brass nut. The float operated needle valve maintained the flow rate into the float chamber equal to that being withdrawn through the siphon. Discharge from each siphon was transmitted to a small sump, from which it was periodically pumped to waste by a small

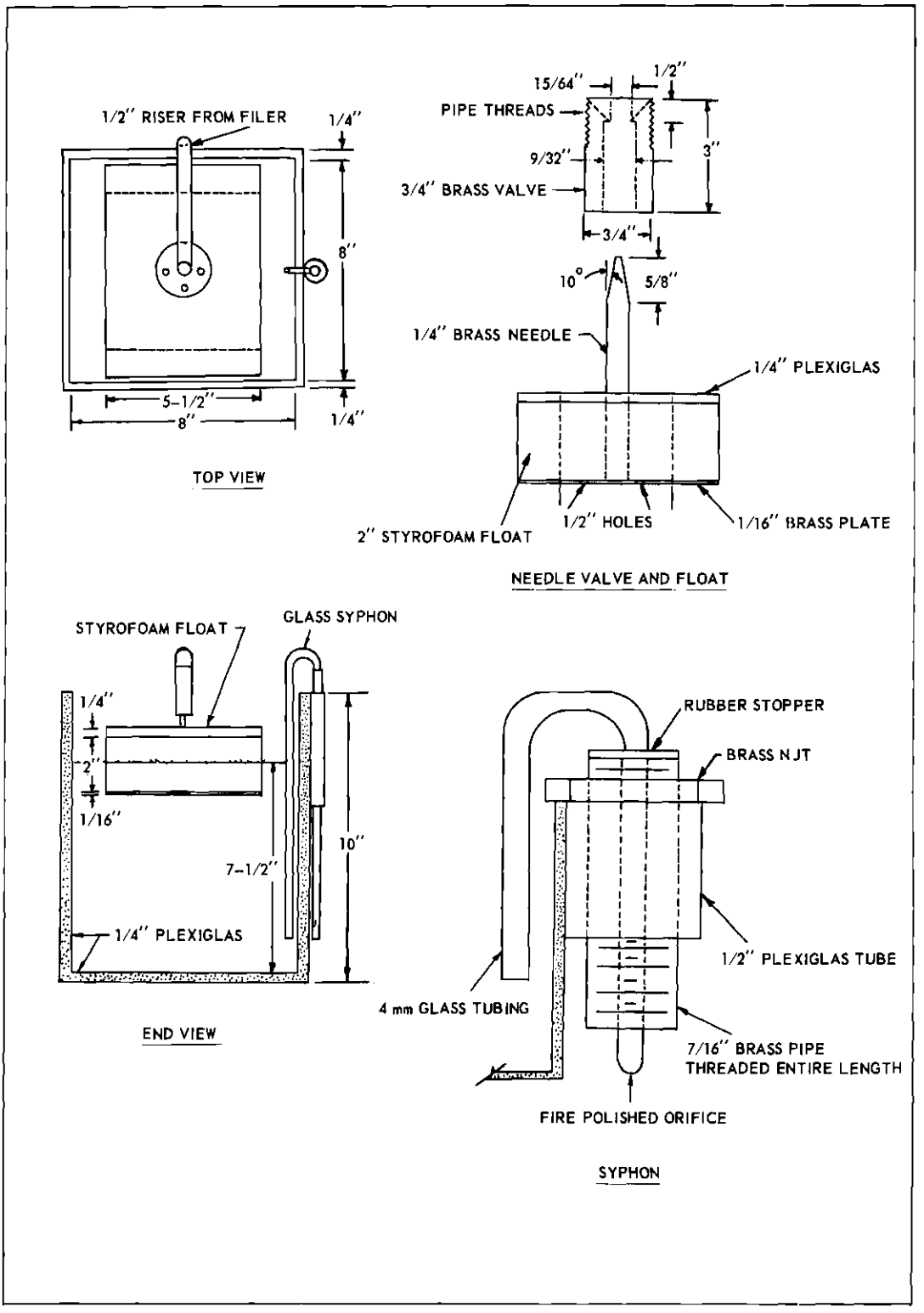


Figure 8. Rate of Flow Controller Details.

float operated sump pump.

Provision was made for backwashing each of the four filters with City of Atlanta tap water. The backwash line was constructed from 1/2 inch galvanized pipe, with branches leading to each filter underdrain. A schematic diagram of the filter piping, including the backwash system, is presented in Figure 9.

It was found that the backwash supply contained significant quantities of air which collected below the screen underdrain during backwashing. These air bubbles produced uneven backwashing and, unless removed, retarded filtration in the region of the bed above the bubble. This condition was successfully corrected by placing an air collection chamber in the backwash line. The chamber was constructed from a two-foot section of 3 inch pipe, capped at each end. Backwash water was introduced into the middle of the chamber and withdrawn near the bottom. Gas in the backwash supply was collected in the top of the chamber, where it was periodically released manually through a valve in the top cap.

The entire apparatus extended vertically through two floors of the laboratory. The piezometric surface above the filters corresponded with the overflow in the constant head tank, a distance of some 11.3 feet above the sand surface and some 14 feet above the filter discharge. In order to facilitate measurement of head loss, it was decided to provide a uniform back pressure on the manometers, thus enabling all of them to be observed on the first floor. This was accomplished through the use of a compressed air manometer suppressor system, a schematic diagram of which is shown in Figure 10.

A manifold was constructed for each filter from 3/4 inch copper

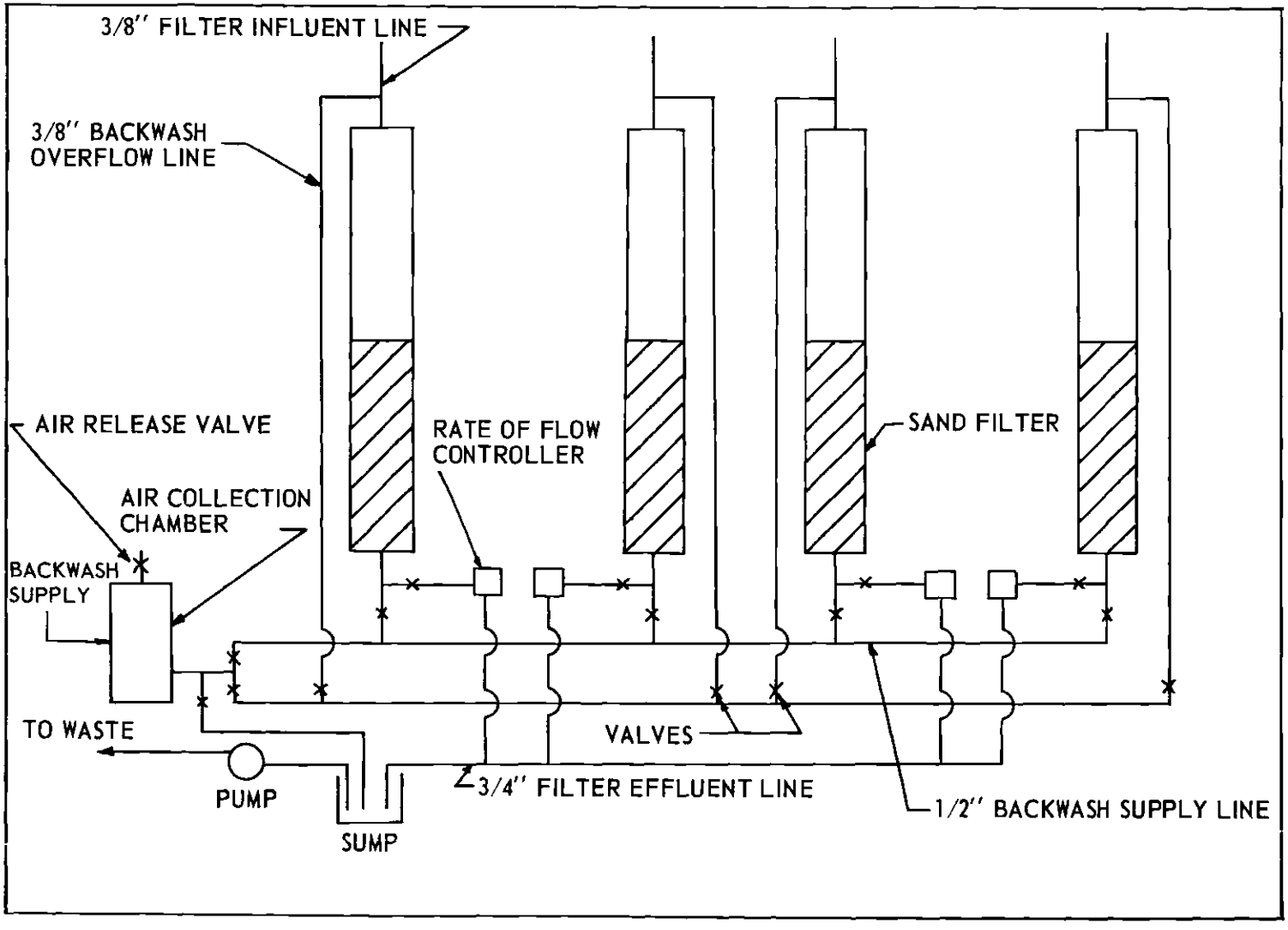


Figure 9. Schematic Diagram of Piping for Filters.

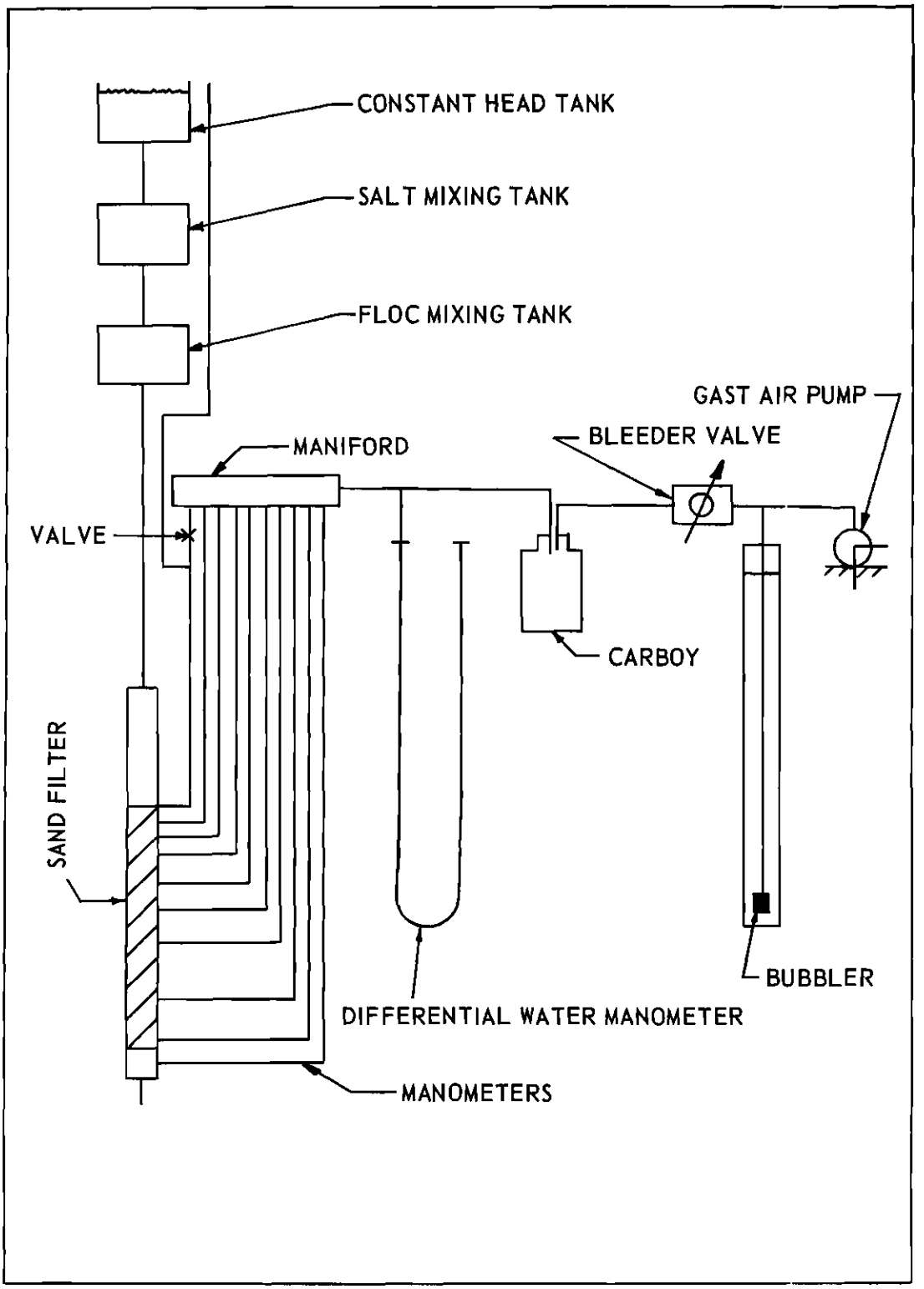


Figure 10. Schematic Diagram of Manometer Suppressor System.

pipe and fitted with ten 1/4 inch ports, one for each manometer connection. The four manifolds were placed in parallel with an air bubbler and connected to the discharge from a Gast air pump. The air pressure applied to the manometers was then controlled at a constant level by the depth of the water column above the air bubbler. A five gallon carboy was placed in the air line leading to the manifolds in order to minimize small fluctuations or pulses in pressure which were produced by the air pump. This precaution prevented surges of flow into and out of the sand bed through the manometers that would have resulted from air pressure fluctuation.

The distance from the sand filter surface to the ceiling of the first floor of the laboratory was about 5.5 feet. When the head loss through a sand filter exceeded this value it was necessary to release some of the air pressure on the manometers. Facilities were provided to accomplish this end. Under these circumstances the top manometer was observed on the second floor at the constant head tank, while the remaining nine manometers could still be observed on the first floor level.

Operation

Considerable preparation was required prior to the start of each filter run. The demineralized water storage tank was filled at least one day in advance of the run. The pH of this water was then adjusted to the desired operating level, and the water allowed to reach room temperature.

Clean Ottawa sand of the -20 + 30 U. S. Standard sieve fraction was placed in each filter on the day preceding each filter run. Sufficient

sand was added so that a bed porosity of 40 per cent could be employed. Based on 3 inch diameter and 24 inch deep filter beds, correcting for the small volume occupied by the sampling points, and using a specific gravity for the sand of 2.64 as measured in the laboratory, it was calculated that 9.67 pounds of sand were required for each filter.

The size of the sand used in this research was determined by three methods as follows:

1. geometric mean of the adjacent sieve openings (0.704 mm),
2. mean width as determined by microscopic measurement of one hundred sand grains (0.802 mm), and
3. Hazen's count and weigh method[91] (0.786 mm).

The latter two procedures gave approximately equal results, but were significantly higher than the value calculated from the rated sieve openings. It is probable that this discrepancy was produced by the use of worn sieves in preparing the sand for use in the filters.

The stock salt solutions and floc suspensions were prepared on the day preceding each run. Floc and salt pumps were calibrated and adjusted to provide the desired pumping rates. Salt pumping rates and stock salt concentrations varied with the aqueous system to be applied to the filter. During the first three runs the floc pumps were set to deliver a flow of 2.50 milliliters per minute. This flow rate was later increased to 5.00 milliliters per minute for the last three runs in order to minimize the possibility of floc settling in the pump discharge lines. The total flow to each filter was maintained at a rate of 372 milliliters per minute (2 gpm per sq ft for a 3 inch diameter filter). The iron concentration in the stock floc suspension was adjusted so that dilution in the floc mix-

ing tank would provide the desired influent iron concentration. For most runs an average iron concentration in the filter influent of 4 mg/l was employed.

Prior to the start of each run the filters were backwashed for a minimum of fifteen minutes. During the backwashing process all air bubbles were removed from manometer lines, sampling points, and from beneath the screen underdrain. A sand bed expansion of fifty per cent was provided.

Backwash flow was terminated gradually allowing the sand to settle to a bed depth of about 25 inches. The backwash overflow line (see Figure 9) was then employed to raise the water level in the filter influent line to a level just below the manometer manifold (5 ft-6 in above the sand surface). The water level was raised at a very slow rate to prevent disturbance of the bed by the flow of water into the manometers. The air lines leading from the manometer manifolds to the suppressor pump were then closed (see Figure 10) to prevent additional water from entering the manometers. The water level was then raised to the level of the effluent lines in the floc mixing tanks. The salt and floc mixing tanks and the constant head tank were then filled with water from the demineralized water storage tank.

After the completion of these preparations, the following procedures were required to place the filters in service.

(1) The filter columns were tapped gently in order to compact the sand beds to a depth of 24 inches and a porosity of 40 per cent.

(2) The water level in the manometer suppressor tube was adjusted to provide the desired amount of suppression. The air pump was then

placed in service and the air flow rate adjusted to provide a small flow rate through the bubbler (see Figure 10). The clamps leading to the filter manometer manifolds were then slowly released and the system allowed to attain equilibrium.

(3) Water from the storage tank was allowed to flow into the constant head tank.

(4) The rate controller siphons were placed in service and adjusted to give the desired flow rate.

(5) The filters were allowed to run for one-half hour to enable the system to reach hydraulic equilibrium.

(6) The salt pump was placed in service, and the filters allowed to run for another half-hour to reach chemical equilibrium.

(7) Manometer readings were obtained to ascertain the head loss across the clean sand beds.

(8) The floc pumps were placed in service.

For a flow rate of 2 gpm per sq ft, the theoretical detention time between the floc pumps and the surface of the bed was 22 minutes. The "zero time" for the run was therefore set as 22 minutes after the floc pumps were started.

The effects of variation in the chemical characteristics of the influent suspension on the filtration process were evaluated in terms of the time rate of filter clogging as measured by head loss increase and by the rate of floc penetration into the filter bed. Head loss increase was determined by hourly observation of the manometer tubes. Rate of floc penetration was evaluated from chemical analyses of samples withdrawn from the beds and from visual observation of the beds. Influent

and effluent samples were collected every two hours and a full series of samples from all sampling points were collected every four hours. Visual observations of bed penetration were made hourly.

During the first two runs the sampling points were allowed to run continuously throughout a run at a flow rate of approximately one milliliter per minute. Subsequent analysis of the data obtained with this procedure led to the conclusion that it was inadequate. The sampling lines required continued adjustment to provide the desired flow rate, a procedure which appeared to remove floc from the bed. Samples obtained during Runs 3, 4, 5, and 6 were obtained by flushing out the sampling tube and then collecting the samples.

The effects of the chemical characteristics of the influent on the ferric floc particles were evaluated in terms of their electrophoretic mobility as determined by microscopic electrophoresis. Influent samples for electrophoretic mobility determinations were withdrawn periodically from the surface sampling point. Additional surface samples were withdrawn from each filter once during a run for microscopic measurement of floc particle size. Effluent samples for determining pH and conductivity were obtained at approximately two hour intervals.

Adjustments to the filter apparatus after the start of a run were only required when head losses in excess of five or six feet were produced. High head loss necessitated a release of the air pressure on the manometer system. As mentioned previously, provision was made for slowly restoring atmospheric pressure to the manometers, necessitating that the sand surface manometer be observed on the second floor. High head loss also produced a slight decrease in flow through the float operated

valve in the rate controller, requiring an adjustment of the siphon. The filters were successfully operated at head losses in excess of 11 feet with only one adjustment required. The maximum reduction in flow rate was less than 10 milliliters per minute (less than 3 per cent) at the time of adjustment.

A filter was removed from service when one of the following occurred:

(1) all of the available head was utilized,

(2) the supply of demineralized water was exhausted, or

(3) the supply of floc was exhausted.

At the termination of a run the filters were backwashed to remove as much floc as possible. The sand was then removed from the filters, washed with 1:1 hydrochloric acid and demineralized water, dried, weighed, and replaced in the filters for the next run.

CHAPTER IV

ANALYTICAL METHODS

General

Analytical methods were employed in this research to characterize the flocculent suspensions, the filter medium, and the filtration process. The following analyses were performed:

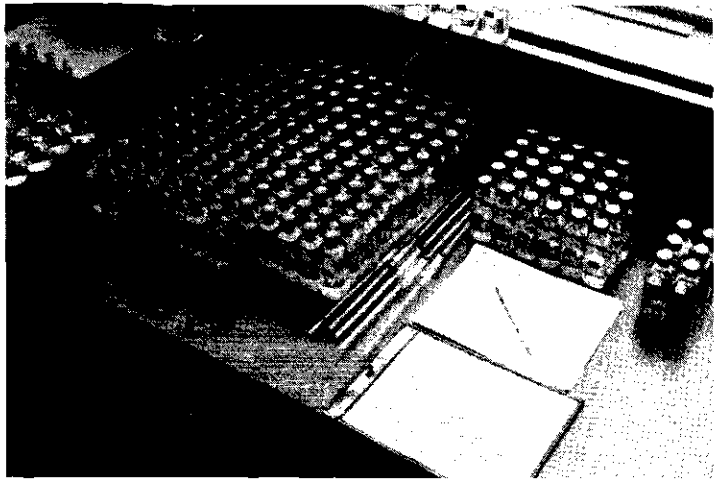
1. analysis for iron concentration,
2. electrophoretic mobility of sand and floc particles,
3. ferric floc particle size,
4. depth of bed penetration,
5. pH determinations
6. specific resistance determinations, and
7. head loss measurement.

Photographs of some of the analytical equipment are presented in Figure 11.

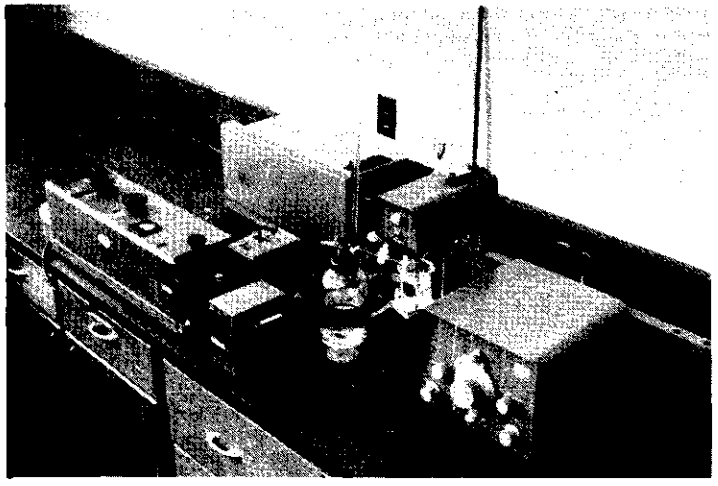
Iron Analysis

Iron coagulant was selected as the particulant material to be investigated. It is representative of the flocculant materials in a typical rapid sand filter influent and can be detected quantitatively with relative ease. Of additional importance to this research, its surface properties vary considerably with the chemical composition of the aqueous phase.

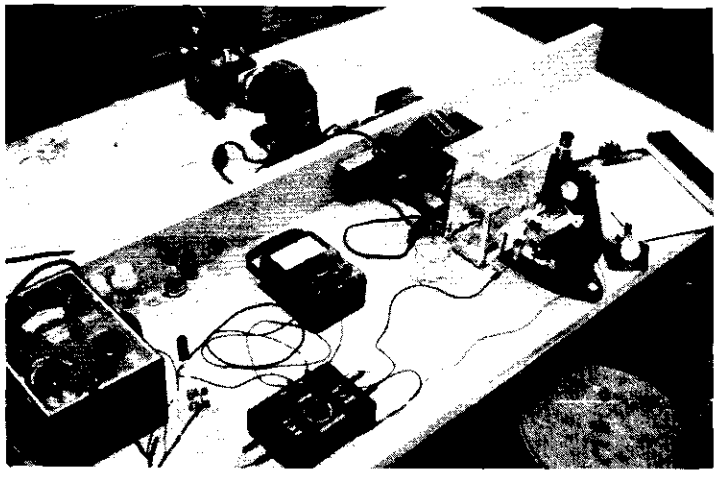
The phenanthroline method as presented in the eleventh edition of



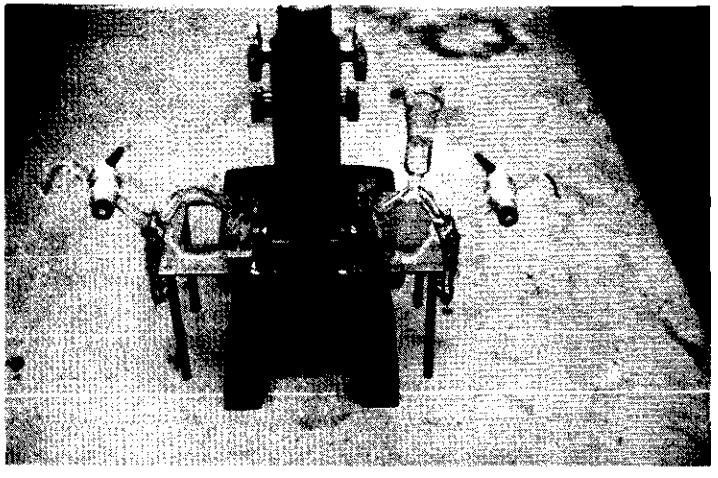
A. Sample Bottles, Glassware, Data Sheets, and Standard Iron Curve.



B. Beckman DU Spectrometer, left; Beckman Model K. Automatic Titrator, right.



C. Microelectrophoresis Equipment, foreground; Particle Sizing Equipment, background.



D. Briggs Cell Mounted on Microscope.

Figure 11. Analytical Equipment.

Standard Methods[92] was modified slightly for analysis of iron concentration. This test is based on the formation of the phenanthroline-ferrous iron complex. Solutions of this complex transmit light in inverse proportion to their iron content in accordance with Beer's law. Digestion and extraction steps necessary when organically bound iron or interferring ions are present were omitted from the analytical procedure. The use of aqueous systems of known composition and prompt acidification of the samples made these steps unnecessary.

The samples were collected in 120 milliliter clear flint glass bottles with plastic caps. The required volume of hydrochloric acid (2 ml/50 ml sample) was added directly to the bottle in which the sample was collected. Difficulty was encountered in getting a representative sample for analysis if the acid was not added immediately. This was due to settling of the ferric hydroxide floc and difficulty in redispersing them uniformly after they had settled.

Higher iron concentrations than those outlined in Standard Methods[92] were used, and it was found convenient to increase the concentration of the phananthrolein reagent. The following amounts and concentration of reagents were added in sequential order to a 50 milliliter sample placed in a 100 ml volumetric flask:

1. Concentrated HCl - 2 ml/50 ml sample added immediately after sample collection.
2. Hydroxylamine reagent - 1 ml (100 g/l hydroxylamine hydrochloride).
3. Ammonium acetate buffer - an amount necessary to produce a final pH of 3.2 to 3.5. This amount varied with each batch of

buffer prepared, and ranged from 12 to 17 ml. (prepared by dissolving 250 g of ammonium acetate in 150 ml demineralized water and diluting to 1000 ml with glacial acetic acid).

4. Phenanthroline reagent - 4 ml (prepared by dissolving 2.5 g 1, 10 - phenantholine monohydrate in one liter of distilled and demineralized water. Solution was facilitated by adding about one ml of concentrated HCl per liter and gently heating).
5. Sufficient distilled and demineralized water was then added to bring the total volume to 100 ml.

The volumetric flasks were shaken after the addition of each chemical and were then inverted 10 times after the addition of the distilled and demineralized water to insure complete mixing. The "color" was allowed to develop for at least 30 minutes before the samples were analyzed.

Standard curves were prepared from a standard solution of ferric iron. Since a new standard curve must be obtained each time a new batch of chemical reagents is prepared, it was deemed advisable to prepare large quantities of reagents. An interesting problem developed from the handling of a large quantity of the ammonium acetate buffer. The prepared buffer was stored in a five gallon carboy from which it was to be syphoned through tygon tubing. It was noted that a white precipitate formed upon addition of the newly prepared buffer to the sample. Rubber tubing was substituted for the tygon but did not remedy the problem. It was finally decided to let the buffer come in contact with glass only and this procedure eliminated the problem. The precipitate was formed when the buffer (after having been in contact with the tubing) was mixed with chloride ions. This was determined by adding HCl and NaCl to the

buffer alone. Both reagents produced the precipitate.

The complexed iron samples were analyzed with a Beckman DU Spectrophometer equiped with a power supply. A Tungsten lamp was employed with a wave length of 510mμ. No filters were employed. Two cell widths were used: 1 cm for samples containing from 0.5 to 5.0 mg/l of iron, and 10 cm for samples containing from 0.00 to 0.50 mg/l.

All glassware in sample collection and analysis was carefully acid washed (with 1:1 HCl) before use. Pipets were also acid washed after each use.

Electrophoretic Mobility Determinations

Microelectrophoretic techniques were employed to determine the mobility of floc and sand particles used in this research. Microelectrophoresis employs the direct observation of microscopically visible particles as they migrate in an electric field. Their speed of migration is determined by measuring the time required for particles to travel a known distance. The electrophoretic mobility[93,94] may then be calculated from the following equation:

$$M = \frac{dX}{tIR_s}$$

where d = distance across which the migration is observed (microns),

X = cross sectional area of the microelectrophoresis cell normal to the direction of particle travel (sq. cm.),

t = time required to traverse the distance d (seconds),

I = current flowing between the cell electrodes (amperes),

R_s = specific resistance of the sample (ohm - cm.), and

M = electrophoretic mobility (microns/sec/volt/cm).

It was necessary to calibrate both the microelectrophoresis cell and the microscope. A Brigg's microelectrophoresis flat cell as reported by Black and Smith[93] was used in conjunction with an American optical monocular microscope equipped with a 30X compensating eyepiece, 10X apochromatic objective, and a dark field condenser. The linearity of the vertical micrometer scale was checked by alternately focusing on the top and bottom of a glass slide. The scale was found to be approximately linear over the entire range but was exactly linear in the range of 200 to 600 microns.

A Whipple ocular micrometer disc was installed in the eyepiece and then calibrated with a stage micrometer. The smallest square was found to measure 28.6 μ with the next larger square measuring 143 μ. These squares were employed in timing the particles.

The cross sectional area of the cell was then determined. The depth was determined by alternately focusing on the top and bottom of the cell in a number of places and then averaging the values determined from the vertical micrometer scale. These measurements were facilitated by placing a small amount of silicic acid in the cell and then blowing the excess out of the cell. A few very small particles remained on the top and bottom of the cell and greatly facilitated focusing. The width of the cell was measured with a graduated mechanical stage. Several measurements were taken at points along the cell and averaged. A methyl orange solution was placed in the cell in order to make its edges more distinct under the microscope. The average depth was 986 μ and the av-

erage width, 21.3 mm, thereby giving a value for X (cross sectional area) of 0.210 square centimeters.

There are only two planes within a microelectrophoresis cell that can be used to measure the true mobility of particles. The two positions are termed the upper and lower stationery layers. When a difference of potential is applied across the cell, a streaming of the fluid takes place. This phenomenon is known as electroendosmosis and occurs as a result of a potential that exists between the liquid and the glass surface of the cell. The liquid flows along the surface of the cell and returns in the center. Thus there are two levels at which the streaming of the fluid does not interfer with the velocity of the particles. Komagata has derived the following relationship[95] from which the vertical distance to the stationary levels can be calculated:

$$a = \frac{1}{2} + \left(\frac{1}{12} + \frac{32}{\pi^5 K} \right)^{\frac{1}{2}}$$

Where a = distance from the top inside surface of a cell of unit depth to the upper stationary layer, and

K = ratio of cell width to cell thickness.

The distance from the top or bottom of the cell to the upper of lower stationary layer can then be calculated from the quantity (1-a) times the average cell depth.

For the cell employed in this research the quantity (1-a) was found to be 0.203. Using an apparent cell depth corrected for the refractive index of water, the stationary layers were calculated to be

150 microns from the top and bottom of the cell. All electrophoretic measurements were made at the lower stationary layer as the dark field condenser could not be focused on the upper layer. The microscope was first focused on the bottom of the cell with the help of a thin line etched on the bottom surface for that purpose. The objective was then raised a distance of 150 microns as determined by the vertical micrometer scale, permitting observation of the lower stationary level.

After the characteristics of the cell and microscope were determined the calibration was checked by measuring the mobility of human red blood cells in $\frac{M}{15}$ phosphate buffer at pH 7.40. A value of - 1.30 μ/sec/volt/cm was determined. Abramson[95] reports the mean value for the mobility of human erythrocytes to be - 1.31 $\pm$0.02 μ/sec/volt/cm. Based on these results, the calibration was accepted.

Mobility samples were analyzed as soon as possible after collection. Ten particles were timed in each direction with the cell polarity being changed after the timing of five particles. The mean value of the twenty observations was used to calculate the mobility. The pH, temperature, and specific resistance of each sample were also recorded. The laboratory manual on microelectrophoresis by Black and Smith[94] was found most helpful. The general procedures, cautions, and corrective measures are outlined in this publication and will not be reproduced here.

Sizing of Floc Particles

Samples for floc particle size measurement were collected from each filter influent once during each run. Microscopic observation of the maximum and minimum demensions of twenty-five particles were made

immediately after the sample collection. Measurements were made employing a 43X objective in conjunction with a filar micrometer equipped with a 12.5X Ramsden eyepiece.

Bed Penetration

The rate of bed penetration has been employed by numerous previous investigators to characterize the filterability of flocculant suspensions in rapid sand filtration. In this research the rate of bed penetration was determined from visual observation of the bed, reinforced by chemical iron analysis of samples withdrawn from within the bed.

The reddish-brown color of the ferric floc retained in the beds was easily distinguished from the clean white Ottawa sand employed as the filter medium. Depth of penetration was determined hourly by direct measurement. It is recognized that this procedure is not as quantitative as might be desired. However, it does provide a simple and valuable means of evaluating the ability of a filter to remove floc from suspension. Comparison of visual filter performance with the results of the chemical analysis of samples withdrawn from the bed was quite good. Figure 2 is a photograph obtained after four hours of elapsed time during Run 6. Bed penetration variations between the filters can easily be observed.

pH Determinations

The pH was determined by a Beckman Zeromatic pH meter equipped with glass and calomel electrodes. Beckman pH 4, 7, and 10 buffers were used to standardize the meter at the proper temperature before each use. This instrument is reported accurate to 0.05 pH units and reproducible

to 0.02 pH units.

Specific Resistance

The specific resistance of all solutions was measured with a model RC-16B2 conductivity bridge manufactured by Industrial Instruments. A conductivity cell with a cell constant of 0.998 was used for all solutions of high conductivity and a cell with a constant of 0.0975 for all solutions of low conducitvity. The conductivity cells were standardized as outlined in Standard Methods[92].

Head Loss

The time rate of filter clogging as measured by head loss increase is a parameter of utmost importance in evaluating the filtration process. It determines the operating time that a filter can be expected to maintain and also provides an indication of the amount and distribution of suspended material which has been retained in the filter bed.

The greatest head loss increase normally occurs in the upper regions of the sand bed, and for this reason manometers were placed at closer intervals near the surface than in the deeper regions of the bed. Increase in head loss was determined from hourly readings of all manometers. Figure 2 depicts the head loss pattern observed after four hours of elapsed time during Run 6.

CHAPTER V

EXPERIMENTAL PROCEDURE

The experimental apparatus was designed to eliminate variation in such physical filter parameters as flow rate (2 gpm/sq ft), bed depth (24 inches), bed porosity (40 per cent) and sand size (-20 +30 U.S. Standard sieve fraction). The experimental procedures were designed to minimize or eliminate variation in the following parameters, most of which are characteristics of the filter influent:

1. influent iron concentration,
2. floc particle size,
3. floc age,
4. temperature,
5. character and structure of the floc, and
6. surface condition of the sand grains at the start of a run.

Concentrated floc suspensions were prepared for each filter on the day preceding each run and contained equal concentrations of ferric iron. A uniform iron concentration in the influent to each filter was obtained by pumping with four identical positive displacement pumps mounted on the same shaft and operated at the same speed by a single motor.

The floc size in the influent to each filter was maintained at reasonably constant levels by providing uniform mixing and pumping facilities from the floc storage tanks through the entire apparatus to the filter surface. The floc particles which reached the filters were found to be

about 20 microns in size.

It was desired to maintain the structure and character of each floc suspension as uniform as possible throughout a filter run. Prior to an experimental run concentrated flocculant suspensions were prepared on a small scale simulating actual filter run conditions. These suspensions were prepared with the same pH, salt type and concentration, and iron concentration as those to be employed during the run. They were allowed to "age" for one day in order to simulate experimental conditions, after which they were diluted to provide the desired filter influent iron concentration. Measurements were then made of the electrophoretic mobility of the ferric floc particles.

These preliminary suspensions were prepared with an automatic titrator (Beckman model K) to eliminate as much variation in floc structure and character as possible during the formation of the floc. The general procedure was as follows:

1. Prepare stock salt solutions.
2. Prepare stock sodium hydroxide solution.
3. Prepare stock ferric chloride solution.
4. Measure out a volume of demineralized water less than the final volume desired providing for addition of salt, ferric chloride and sodium hydroxide solutions.
5. Add sufficient salt solution to provide the desired salt concentration in the final volume.
6. Set the desired pH on the control of the automatic titrator.
7. Slowly add ferric chloride manually allowing the automatic titrator to maintain the desired pH by adding sodium hydroxide.

(The solution is adjusted initially to the desired pH by addition of either sodium hydroxide or ferric chloride.)

8. After sufficient ferric chloride was added to provide the desired final concentration of iron, demineralized water was added to bring the suspension to the final desired volume. (With trial the amount of demineralized water could be kept to a minimum. In any case the amount was so small that it had no significant affect on pH.)
9. The concentrated flocculent suspensions were allowed to age for one day.
10. Dilute floc suspensions were prepared by mixing demineralized water, stock salt solution, and concentrated floc suspension to produce aqueous suspensions containing salt and iron concentrations equal to those to be employed during the filter run.
11. The electrophoretic mobility and pH of the dilute floc suspensions were determined.

It should be noted that the floc suspensions were prepared and maintained under conditions which were equivalent to those employed during the experimental runs.

Ferric floc suspensions for the experimental runs were prepared in a similar manner with the exception that the automatic titrator was not employed. A pH meter was substituted for the titrator since it was not considered adaptable to preparation of the larger volumes required. Both the ferric chloride and sodium hydroxide were slowly added manually in order to maintain a constant pH as indicated by the meter. After

preparation, the floc was placed in storage tanks where it was mixed prior to and throughout the experimental run.

The sand was carefully washed with 1:1 hydrochloric acid between runs so that the surface condition of the grains would be identical for each run. The sand was rinsed with demineralized water, dried at 100° C, weighed, and replaced in the filter column. The temperature of the suspending medium was maintained as constant as possible by allowing it to come to equilibrium with the room temperature prior to the start of each run.

The first two experimental runs were to check out the operation of the model filter apparatus and to see if the results were reproducible between the filters and also between runs. Tap water containing 50 mg/l of added sodium sulfate was employed. A ferric floc suspension was prepared and placed in a large mouth jar (5 gallon capacity) from which it could be pumped to each filter. The suspension was mixed during these two runs with a propeller type mixer.

Run 3 was designed to observe possible differences in filtration occasioned by the addition of selected inorganic salts which would in turn affect the electrophoretic mobility of the suspended floc particles. The salts chosen were sodium chloride, sodium sulfate, trisodium phosphate, and magnesium chloride. Aqueous floc suspensions containing one of these salts at a concentration of 50 mg/l in demineralized water were applied to each of the four filters. The pH varied from filter to filter, depending upon the salt used.

The results of the first three runs indicated that the apparatus functioned extremely well, and also provided information for the design

of the next three experiments. It was decided to investigate the effects of selected inorganic anions on the filtration process. Chloride, sulfate, and phosphate anions were selected and their effects at a concentration level of 25 mg/l were observed at pH levels of 5.0, 7.0, and 9.5. Comparison was made with filters treating floc prepared and maintained in demineralized water adjusted to the desired pH level.

A pH of 5.0 was employed for all filters during Run 4. The 500 gallon demineralized water storage tank was adjusted to this pH by addition of hydrochloric acid. Twenty-five mg/l of chloride, sulfate, and phosphate ions (prepared from sodium chloride, sodium sulfate, and monobasic sodium phosphate) were added to three experimental filters. One filter received only pH adjusted demineralized water.

Run 5 was made at a pH of 9.5. Twenty-five mg/l of bicarbonate ion (from sodium bicarbonate) were added to the storage tank to provide some buffer capacity. The tank was then adjusted to pH 9.5 with sodium hydroxide. Twenty-five mg/l of chloride, sulfate, and phosphate (prepared from sodium chloride, sodium sulfate and dibasic sodium phosphate) were added to three of the filters with the other filter again acting as a control.

Run 6 was made at a pH of 7.0. The storage tank was titrated to pH 7.0 with sodium bicarbonate. It was necessary to titrate monobasic sodium phosphate to pH 7.0 with sodium hydroxide for pH control in the phosphate system. Twenty-five mg/l of chloride, sulfate, and phosphate (prepared from sodium chloride, sodium sulfate, and the titrated monobasic sodium phosphate) were then pumped to the filters as before. Again, one filter received only the buffered demineralized water.

It should be noted that a separate flocculant suspension was prepared for each filter for Runs 3 through 6. Each concentrated suspension contained the same added salt concentration and pH level as the aqueous system with which it was mixed.

CHAPTER VI

EXPERIMENTAL RESULTS

The effects of selected inorganic ions on the filtration process have been evaluated in terms of two common criteria for filter performance: (1) time rate of clogging as measured by head loss increase, and (2) the rate of bed penetration.

Six experimental runs were made employing four filters during each run. The first run was designed to check the operation of the model filter apparatus and to check the results for reproducibility between each filter. Tap water containing 50 mg/l of added sodium sulfate was employed. A single concentrated floc suspension provided floc for all four filters. The floc pumps were each operated at a flow rate of 2.5 ml/min to produce an iron concentration in the influent to each filter of 1 mg/l. After 17 hours of operation a total head loss of 1.5 feet and a bed penetration of approximately 15 inches were produced. Three of the filters correlated very closely with one filter deviating slightly. The results were considered to confirm the ability of the experimental apparatus and techniques to yield reproducible data between filters during a run.

Run 2 was designed to check reproducibility of results between filter runs. New sand was placed in two of the filters and sand from Run 1 left in the other two units. The old sand beds were thoroughly backwashed between runs. With this exception, Run 2 was conducted just as Run 1. The two filters containing new sand produced essentially the same results

as Run 1. After 17 hours the penetration for the filters with new sand was approximately 15 inches as before, but only 9 inches in the two filters containing old sand. In addition, a somewhat greater head loss was produced in the filters containing old sand. The electrophoretic mobility of the influent floc particles was observed to be -1.22 μ/sec/volt/cm.

Based on the results of Runs 1 and 2 it was decided that reproducibility between filters and between filter runs was within acceptable limits. Based on the results of Run 2 it was decided to treat all sand with 1:1 hydrochloric acid before reuse in the filters, so that variation in the sand surface characteristics would not complicate the interpretation of results.

Run 3 was designed to test whether differences in filtration could be demonstrated by the addition of selected inorganic salts. Fifty mg/l of NaCl, Na_2SO_4, Na_3PO_4, and $MgCl_2$ were chosen for study. Demineralized water was used in the storage tank for the first time. The influent iron concentration was raised to 4 mg/l in an attempt to minimize variations in influent iron concentrations noted in Runs 1 and 2. It was decided that this increased iron concentration would also produce greater variation in head loss and bed penetration if chemical aspects were indeed important. Table 1 is a summary of the filter influent characteristics employed during Run 3. The observed head loss and bed penetration are plotted against filtration time in Figure 12. Significant differences in filtration can be noted for the four systems studied.

Iron data from samples obtained within the filter beds were not conclusive and in many cases gave concentrations as high as five times the influent iron concentration. This was also noted in Runs 1 and 2,

Table 1. Filter Influent Characteristics for Run 3

	Filter 1	Filter 2	Filter 3	Filter 4
Salt Added	Na_3PO_4	$MgCl_2$	Na_2SO_4	NaCl
Salt Concentration, Mg/l	0 *50 mg/l	50 mg/l	50 mg/l	50 mg/l
pH	6.5 *10.2	6.2	6.5	6.5
Floc Mobility, μ/sec/volt/cm	-1.36 *-1.99	+1.43	-1.29	-1.32
Specific Resistance, Ohm-cm	28,200 *7,650	6,450	8,300	7,300
Average Temperature, ^{o}C	26.0	26.0	26.0	26.0

*Note: Salt pump for Filter 1 operated for only 1 hour during the run. Starred values were observed while the pump was operating; unstarred values were observed during the remainder of the run.

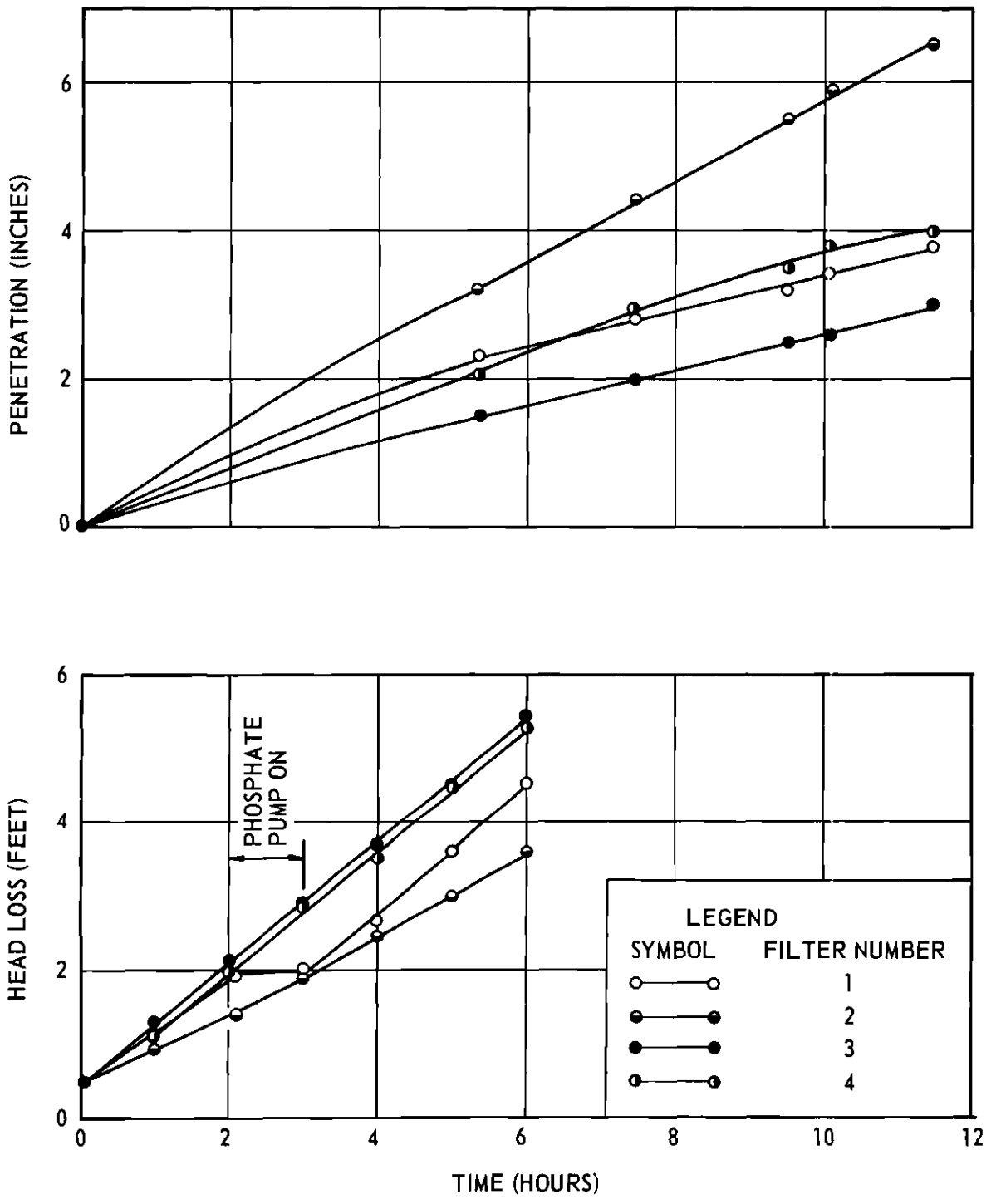


Figure 12. Bed Penetration and Head Loss Versus Time During Run 3.

with the result that the sampling procedure was changed for Run 3. During the initial two runs the sampling points were allowed to flow continuously at a rate of approximately one ml/min, the procedure employed by O'Melia[4]. A great deal of difficulty was encountered in maintaining this low flow rate and frequent adjustments were required. In Run 3 the sampling points were allowed to flow at a rate of about 50 ml/min for a few seconds, after which samples were collected at the flow rate of about one ml/min. This procedure was similar to that employed by Eliassen[88]. The results were again quite erratic, indicating that floc were periodically sheared off by hydrodynamic forces. Samples from within the filter bed were eliminated during all runs made after Run 3. Filters were still sampled to measure bed penetration at the termination of succeeding runs.

Runs 4, 5, and 6 were designed to test the effects of chloride, sulfate, and phosphate ions on the filtration process at different hydrogen ion concentrations. These runs were conducted in a manner similar to Run 3 with one exception: the floc pumping rate was increased to 5.0 ml/min to minimize settling of the floc in the lines leading from the floc storage tanks. These three runs will be discussed collectively. The relationships observed between head loss and bed penetration with the weight of iron applied to the filters are presented in Figures 13, 14, and 15. Filter influent characteristics are presented in Tables 2, 3, and 4. To facilitate a comparison of the effects of a change in hydrogen ion concentration on each aqueous system, Figures 16 through 19 are presented. These and other data presented in this chapter are discussed in detail in Chapter VII.

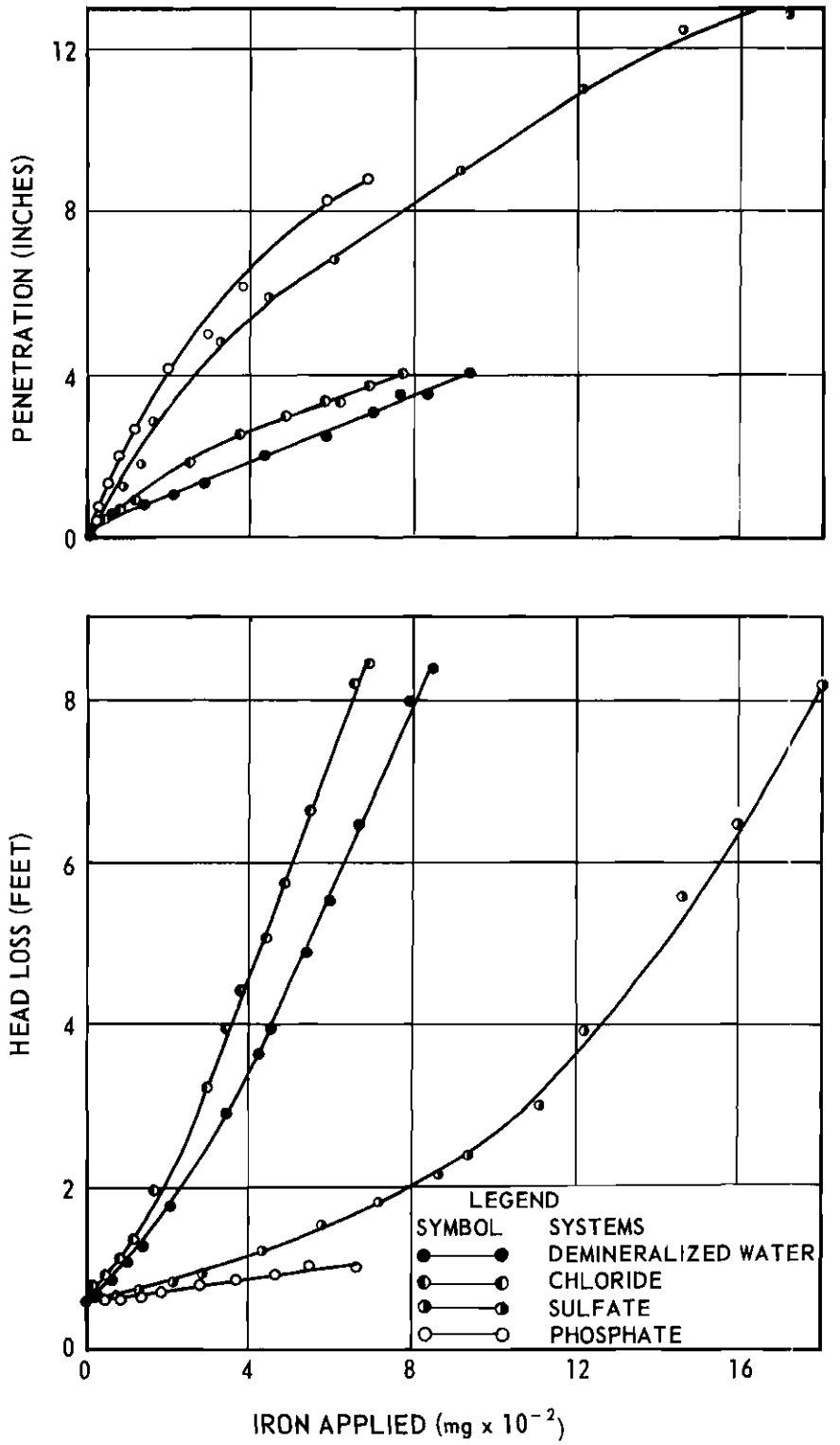


Figure 13. Bed Penetration and Head Loss Versus Total Iron Appled at pH 5.0 During Run 4.

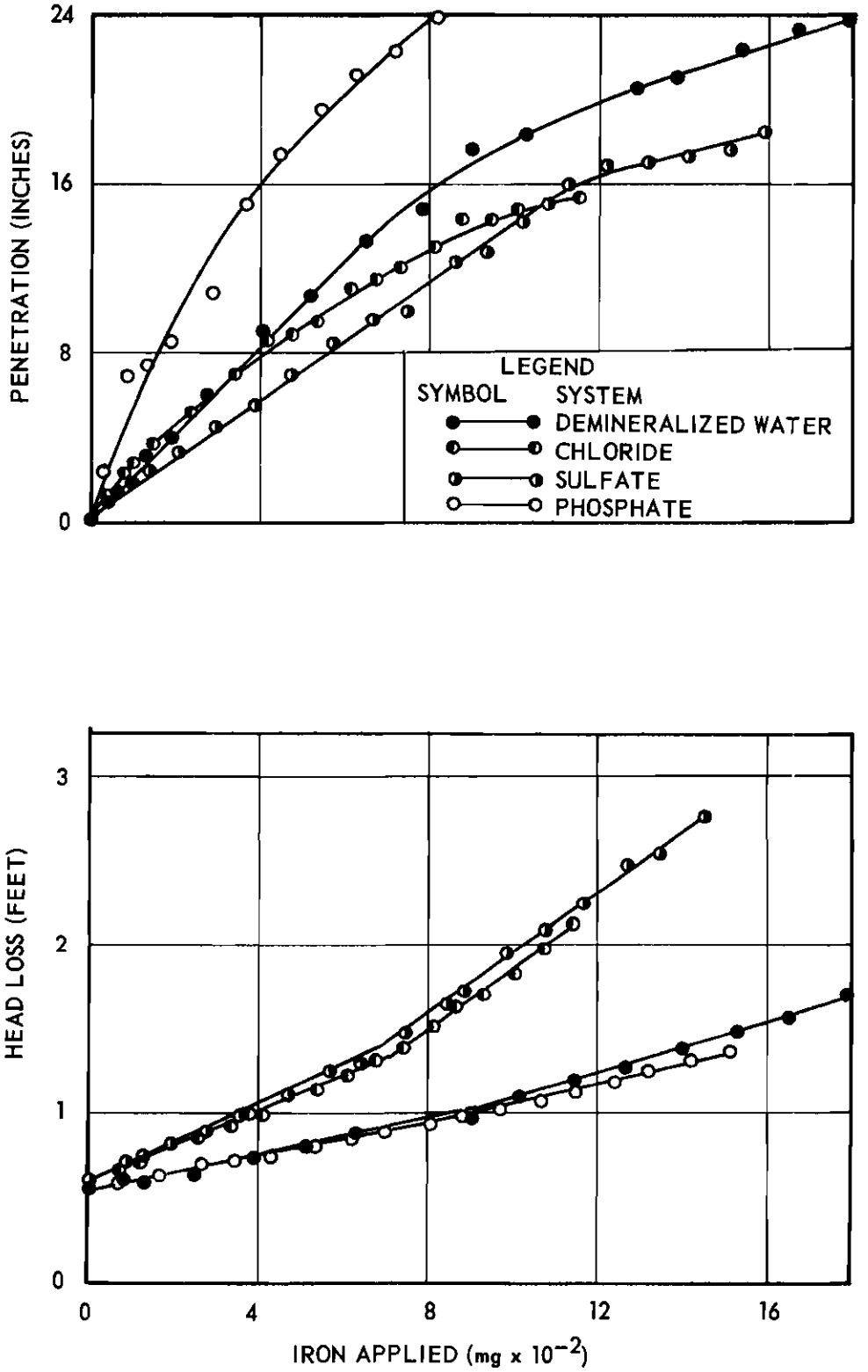


Figure 14. Bed Penetration and Head Loss Versus Total Iron Applied at pH 9.5 During Run 5.

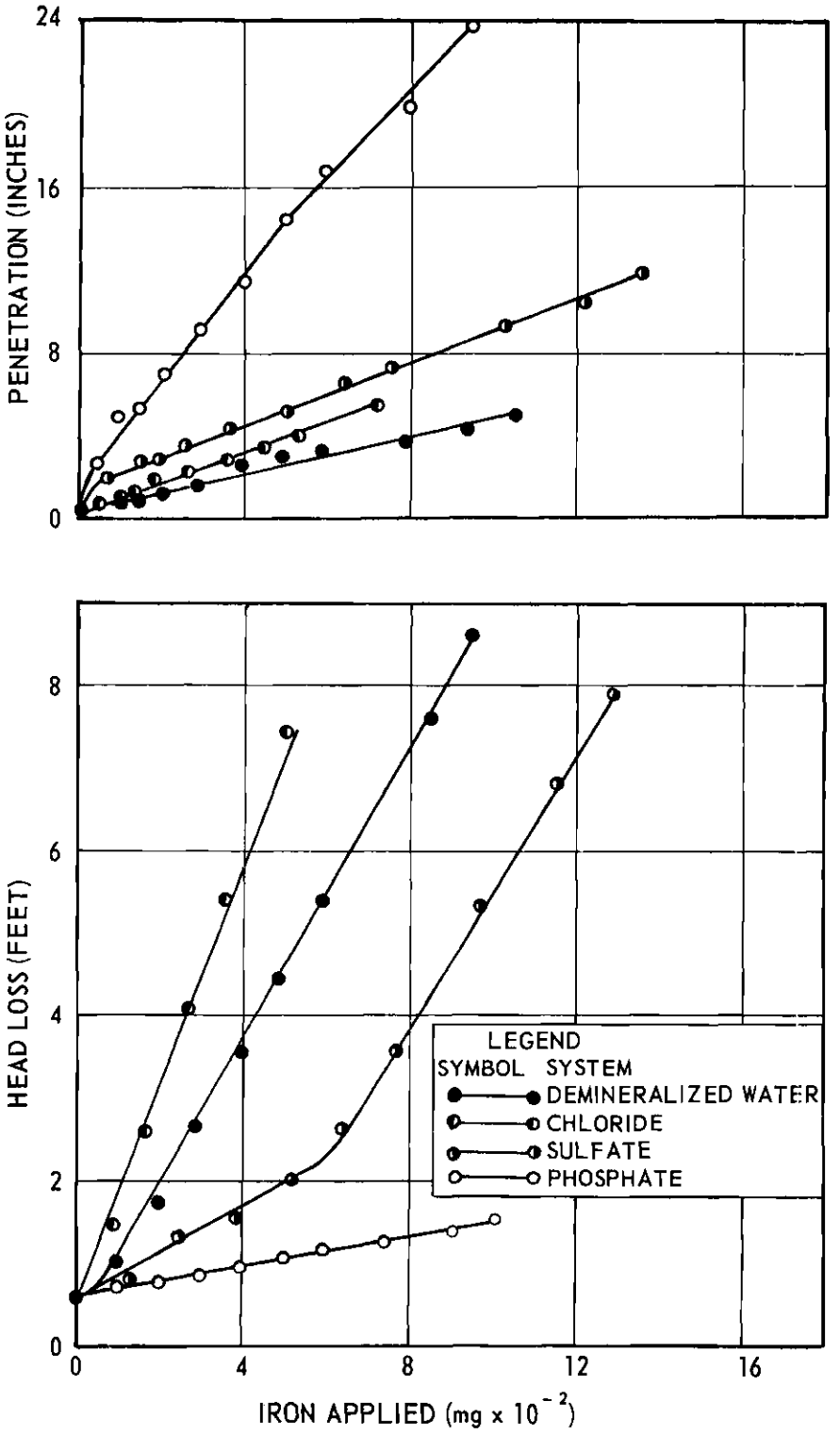


Figure 15. Bed Penetration and Head Loss Versus Total Iron Applied at pH 7.0 During Run 6.

Table 2. Filter Influent Characteristics for Run 4

	Filter 1	Filter 2	Filter 3	Filter 4
Salt Added	None	NaCl	Na_2SO_4	NaH_2PO_4
Salt Concentration, mg/l	0	25 chloride	25 sulfate	25 phosphate
pH	5.0	5.0	5.0	5.0
Floc Mobility, μ/sec/volt/cm	+1.95	+2.23	+0.52	-1.28
Specific Resistance Ohm-cm	28,200	7,600	9,600	19,000
Floc size, μ (long and short dimensions)	20 x 15	18 x 15	21 x 15	15 x 5
Average Temperature, ^{o}C	22.2	22.2	22.2	22.2
pH control: chemical added to storage tank	HCl	HCl	HCl	HCl

Table 3. Filter Influent Characteristics for Run 5

	Filter 1	Filter 2	Filter 3	Filter 4
Salt Added	None	NaCl	Na_2SO_4	Na_2HPO_4
Salt Concentration, mg/l	0	25 chloride	25 sulfate	25 phosphate
pH	9.5	9.5	9.5	9.5
Floc Mobility, μ/sec/volt/cm	-1.72	-1.03	-1.55	-2.16
Specific Resistance, Ohm-cm	8,700	7,800	7,100	8,200
Floc Size, μ (long and short dimensions)	23 x 17	26 x 19	21 x 14	21 x 15
Average Temperature ^{o}C	22.5	22.5	22.5	22.5
pH control: chemicals added to storage tank	25 mg/l HCO_3^- from $NaHCO_3$ + NaOH to pH 9.5	25 mg/l HCO_3^- from $NaHCO_3$ + NaOH to pH 9.5	25 mg/l HCO_3^- from $NaHCO_3$ + NaOH to pH 9.5	25 mg/l HCO_3^- from $NaHCO_3$ + NaOH to pH 9.5

Table 4. Filter Influent Characteristics for Run 6

	Filter 1	Filter 2	Filter 3	Filter 4
Salt Added	None	NaCl	Na_2SO_4	NaH_2PO_4
Salt Concentration, mg/l	0	25 chloride	25 sulfate	25 phosphate
pH	7.0	7.0	7.0	7.0
Floc Mobility, μ/sec/volt/cm	-0.71	-0.64	-0.77	-1.85
Specific Resistance Ohm-cm	29,500	9,900	7,900	13,900
Floc size, μ (long and short dimensions)	26 x 17	15 x 10	16 x 10	20 x 13
Average Temperature, ^{o}C	22.0	22.0	22.0	22.0
pH control: chemical added to storage tank	$NaHCO_3$ to pH 7.0	$NaHCO_3$ to pH 7.0	$NaHCO_3$ to pH 7.0	$NaHCO_3$ to pH 7.0
Chemical added to salt solution	None	None	None	NaOH to pH 7.0

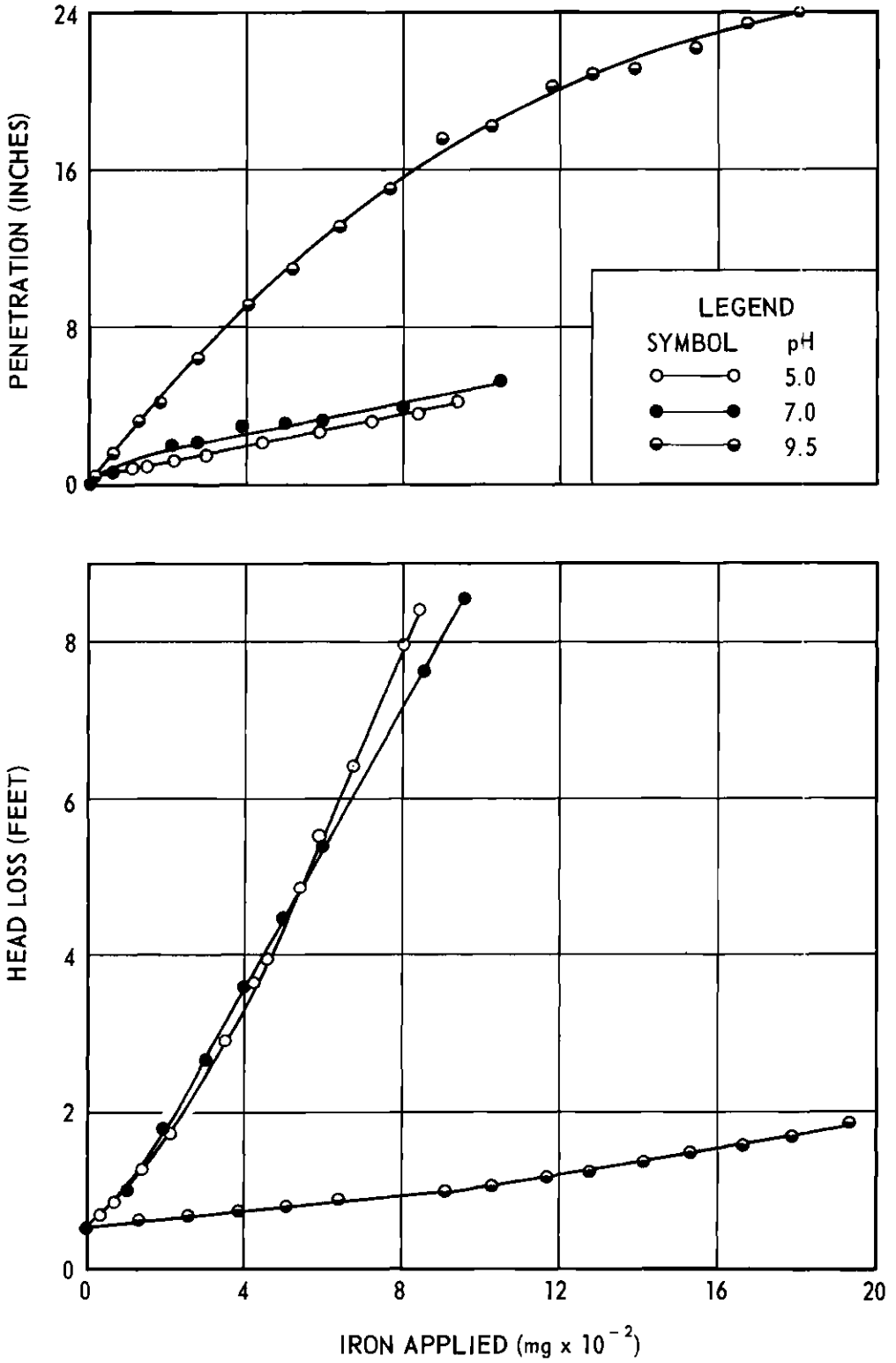


Figure 16. Bed Penetration and Head Loss Versus Total Iron Applied with Demineralized Water.

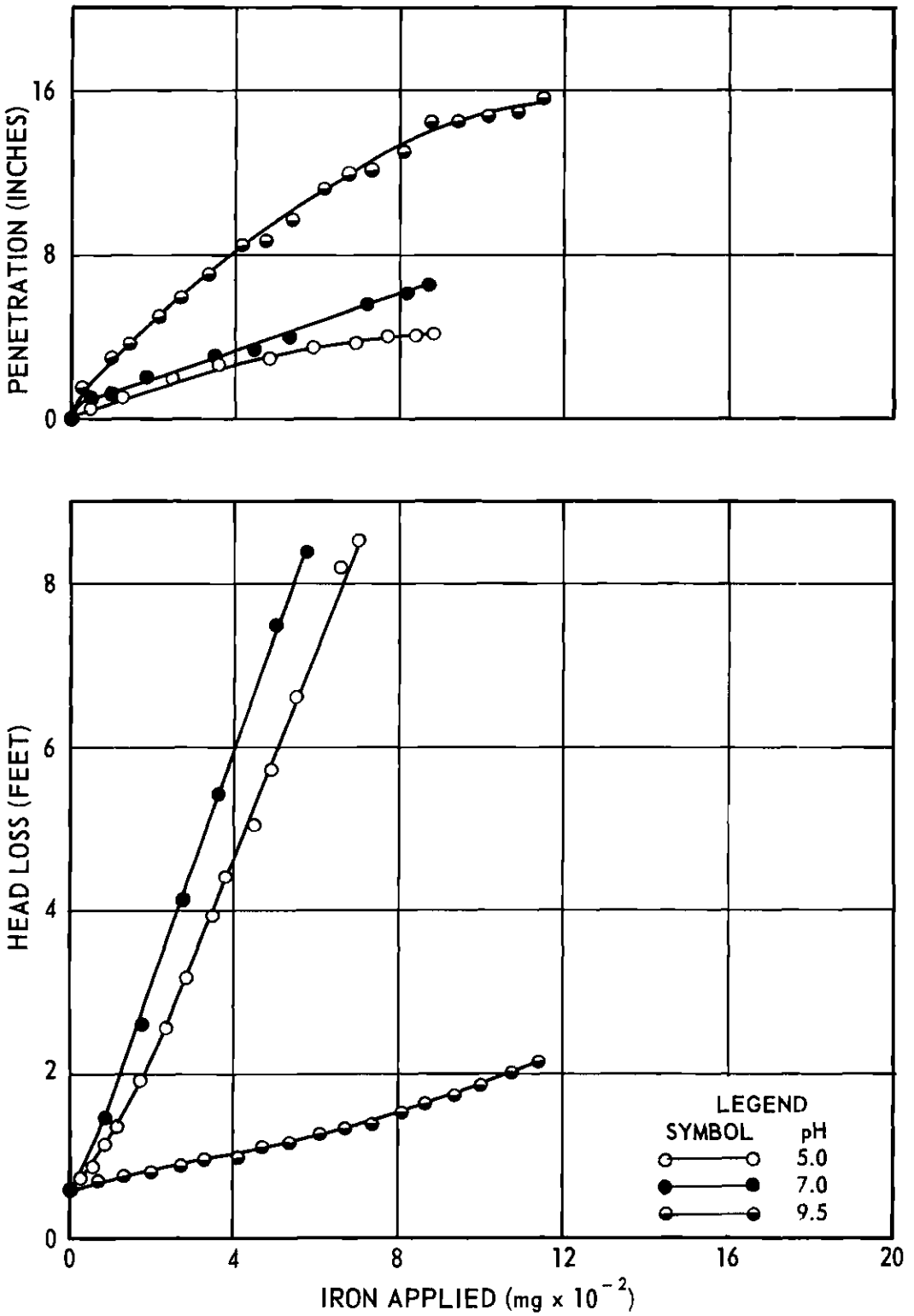


Figure 17. Bed Penetration and Head Loss Versus Total Iron Applied with 25 mg/l Chloride.

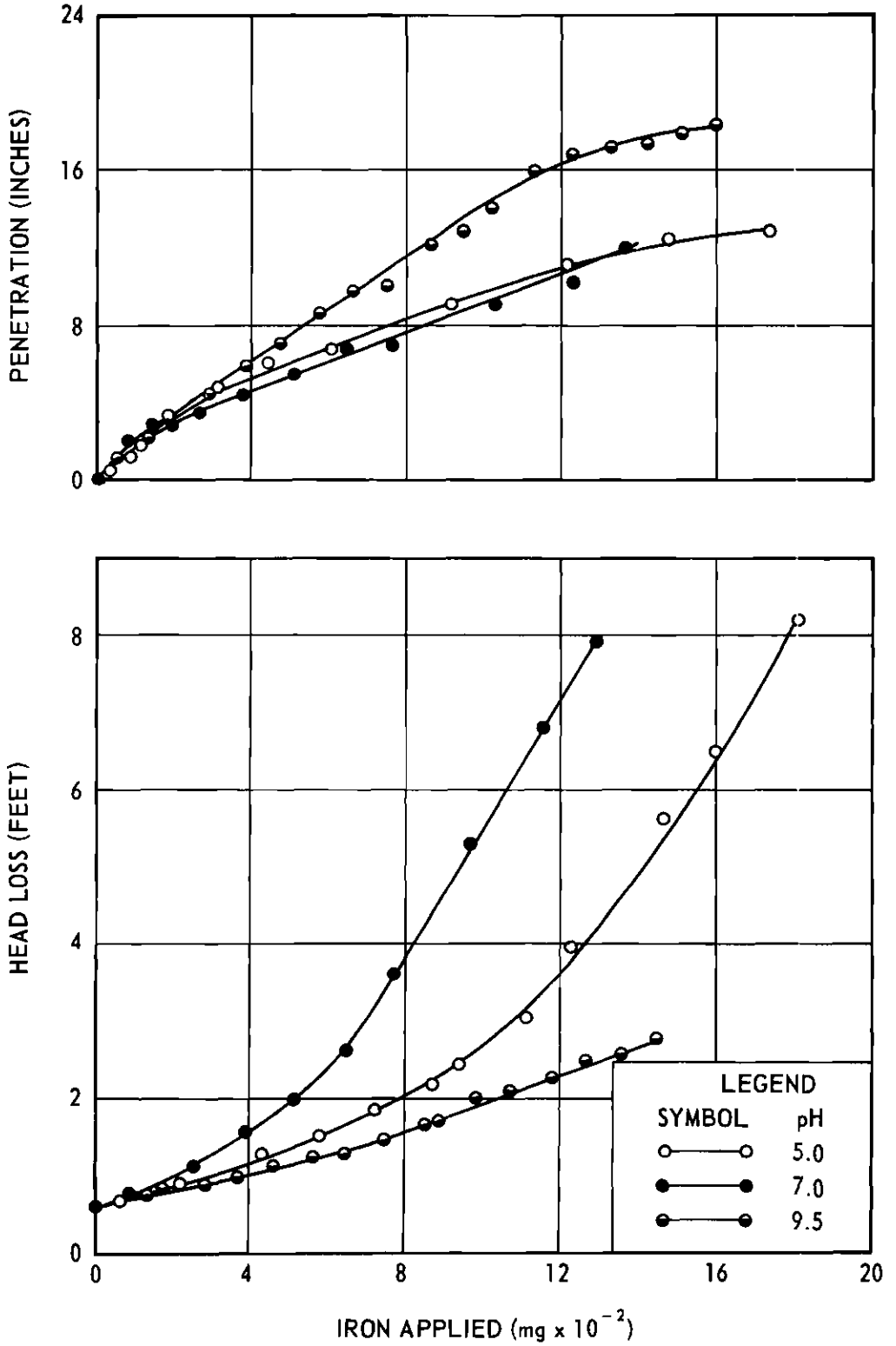


Figure 18. Bed Penetration and Head Loss Versus Total Iron Applied with 25 mg/l Sulfate.

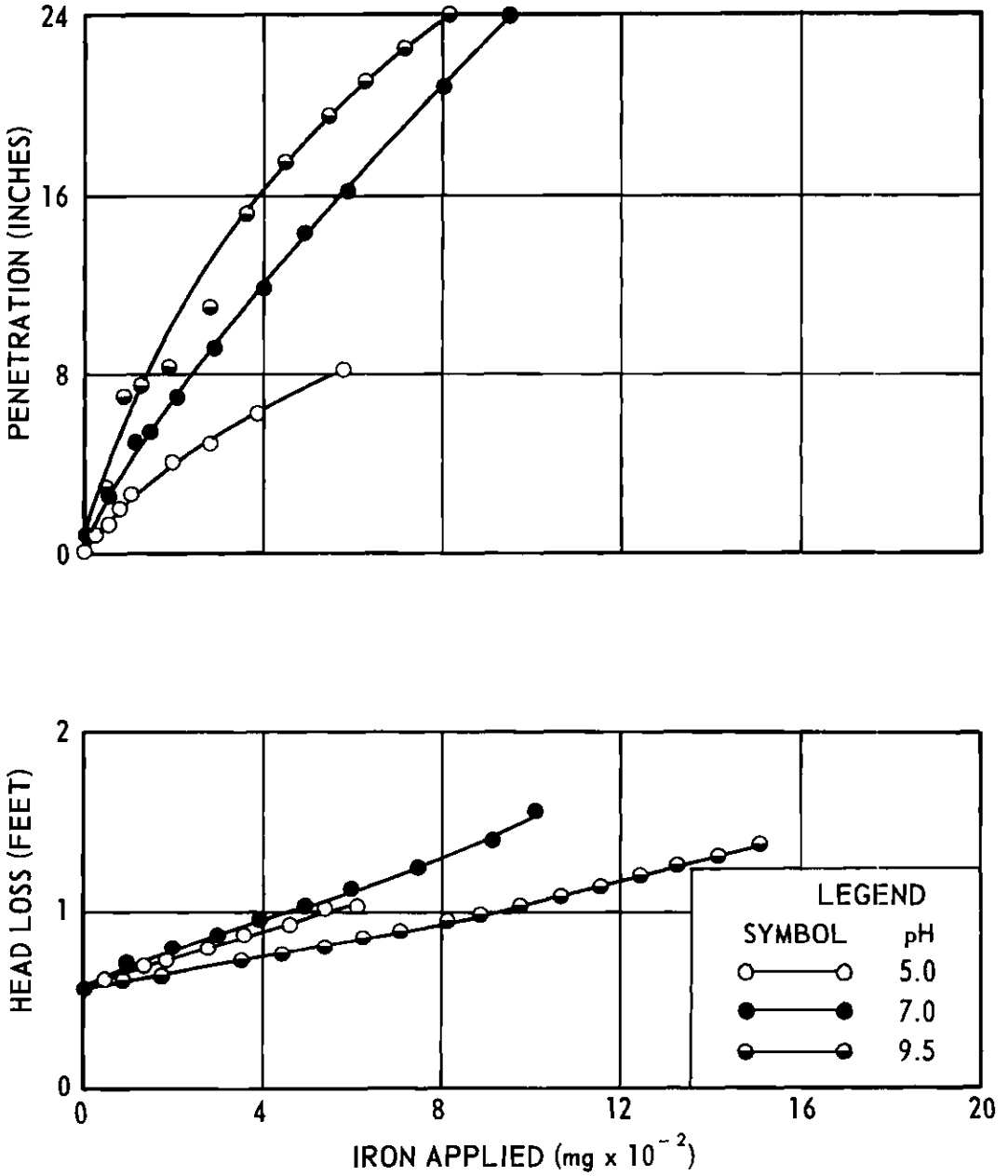


Figure 19. Bed Penetration and Head Loss Versus Total Iron Applied with 25 mg/l Phosphate.

After 7.25 hours of filtration in Run 4 the phosphate salt pump was shut off. The effects resulting from this procedure should indicate the relative importance of ions incorporated into the floc structure and those dissolved in the aqueous phase of the suspension. Figures 20 and 21 present the results of this experiment.

Head loss and bed penetration are both important in the filtration process. Some investigators have advocated the use of very shallow filters since they found the removal of suspended matter to occur very near the surface. A filter that functioned in this manner was observed during Run 6 and is depicted in Figure 22. In contrast to removal at or near the surface, complete bed penetration can result with the removal load being distributed throughout the bed depth. Figure 23 depicts a filtration system of this type observed during Run 5. Figure 24 illustrates an unusual case that was observed during Run 5. It is unique in that there was less head loss across the first inch than the next two inches.

Loss of head is linearly distributed across a sand bed when clean water is passed through it. As suspended matter is removed by the filter this linear relation disappears. The data presented in Figure 25 illustrates this phenomenon. The penetration curve for this filter is superimposed on the head loss curves. Figure 26 contains the same data shown in Figure 25, but presents it in a somewhat different manner. The change in pressure at different depths is more clearly presented by a curve of this type.

After 6.5 hours of operation during Run 4 the filter bed was disturbed due to a malfunction of the manometer suppressor system. This

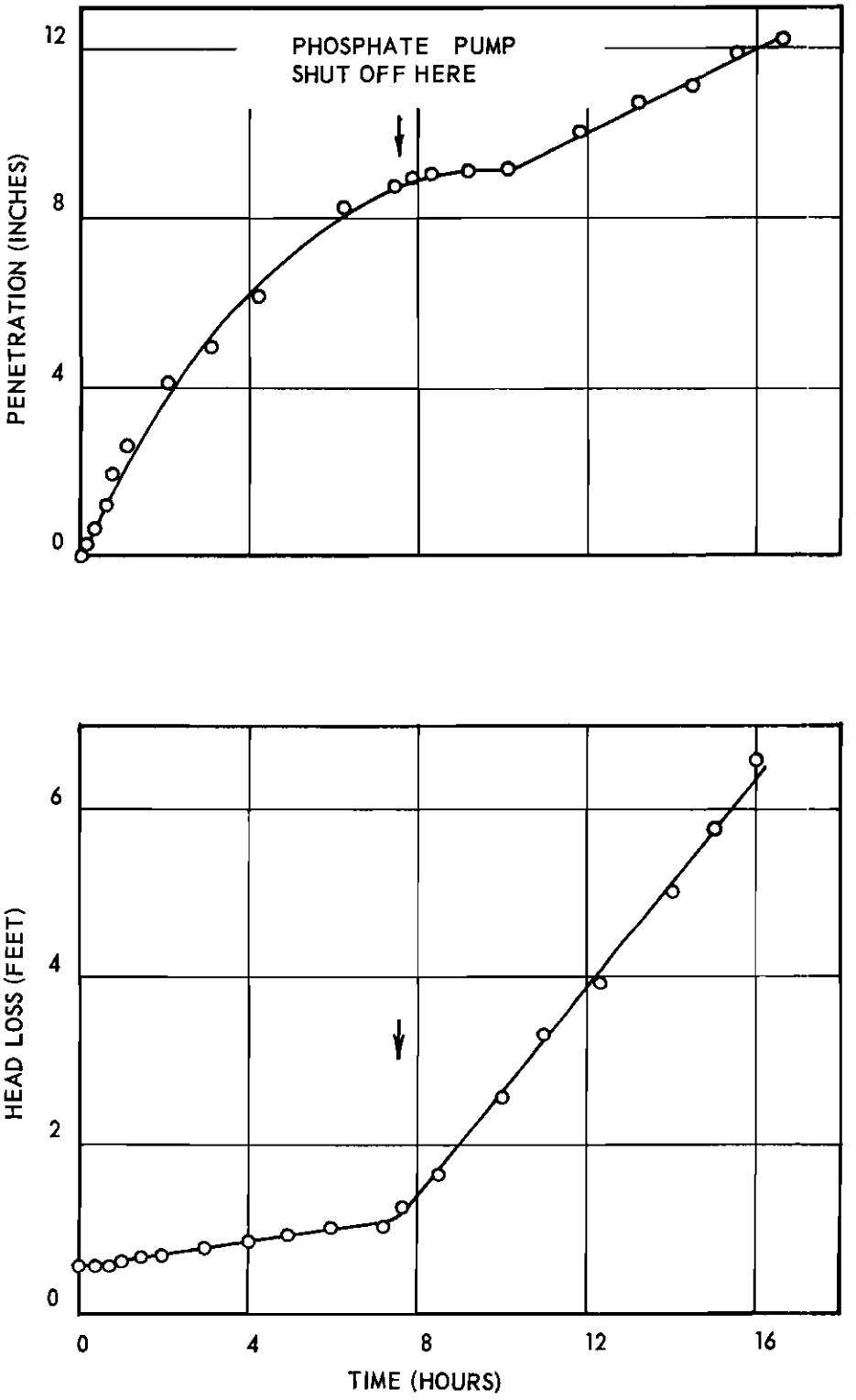


Figure 20. Bed Penetration and Head Loss Versus Time with 25 mg/l Phosphate at pH 5.0 During Run 4.

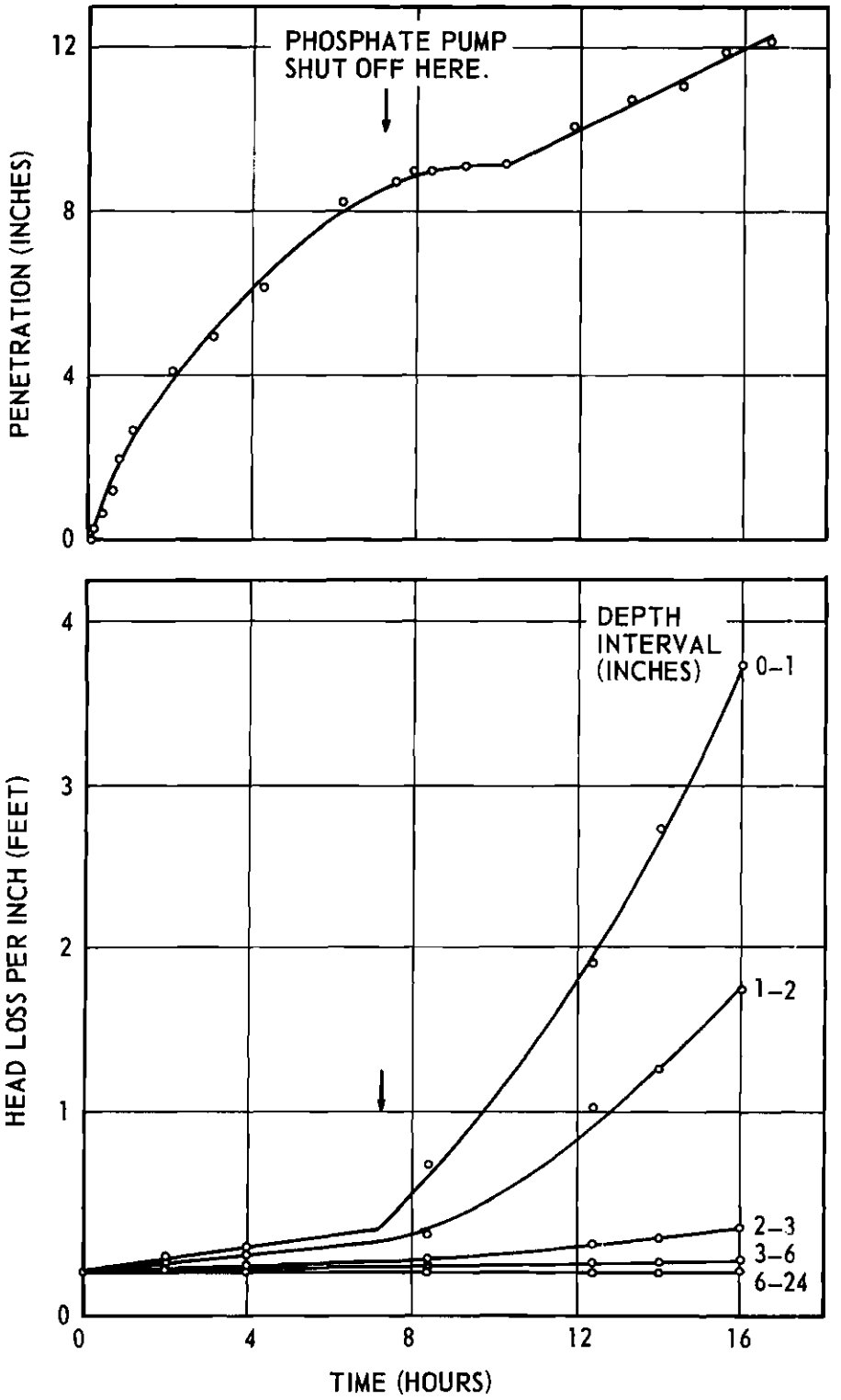


Figure 21. Bed Penetration and Head Loss per Inch Versus Time with 25 mg/l Phosphate at pH 5.0 During Run 4.

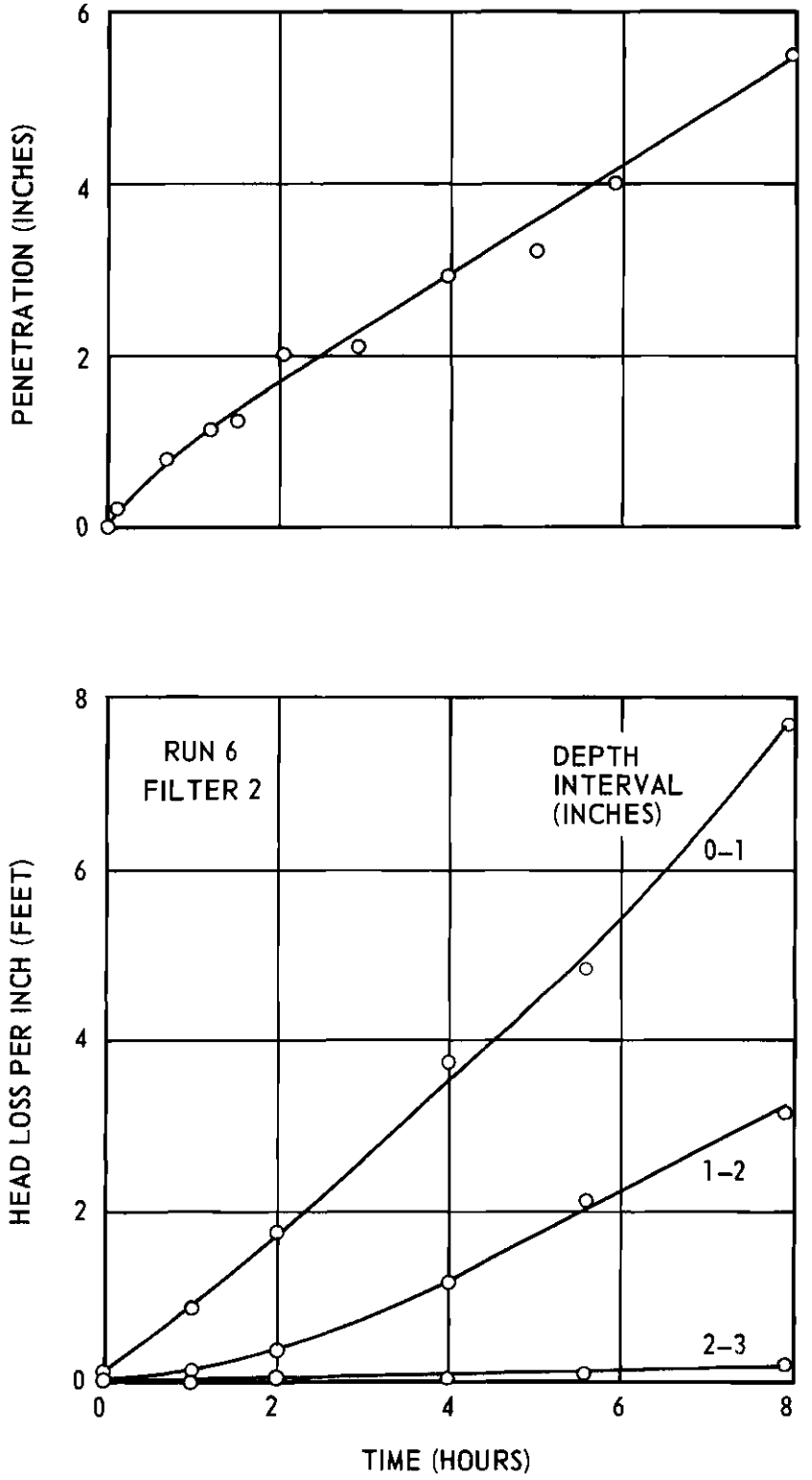


Figure 22. Bed Penetration and Head Loss per Inch for Filter with Removal Near Surface of Filter.

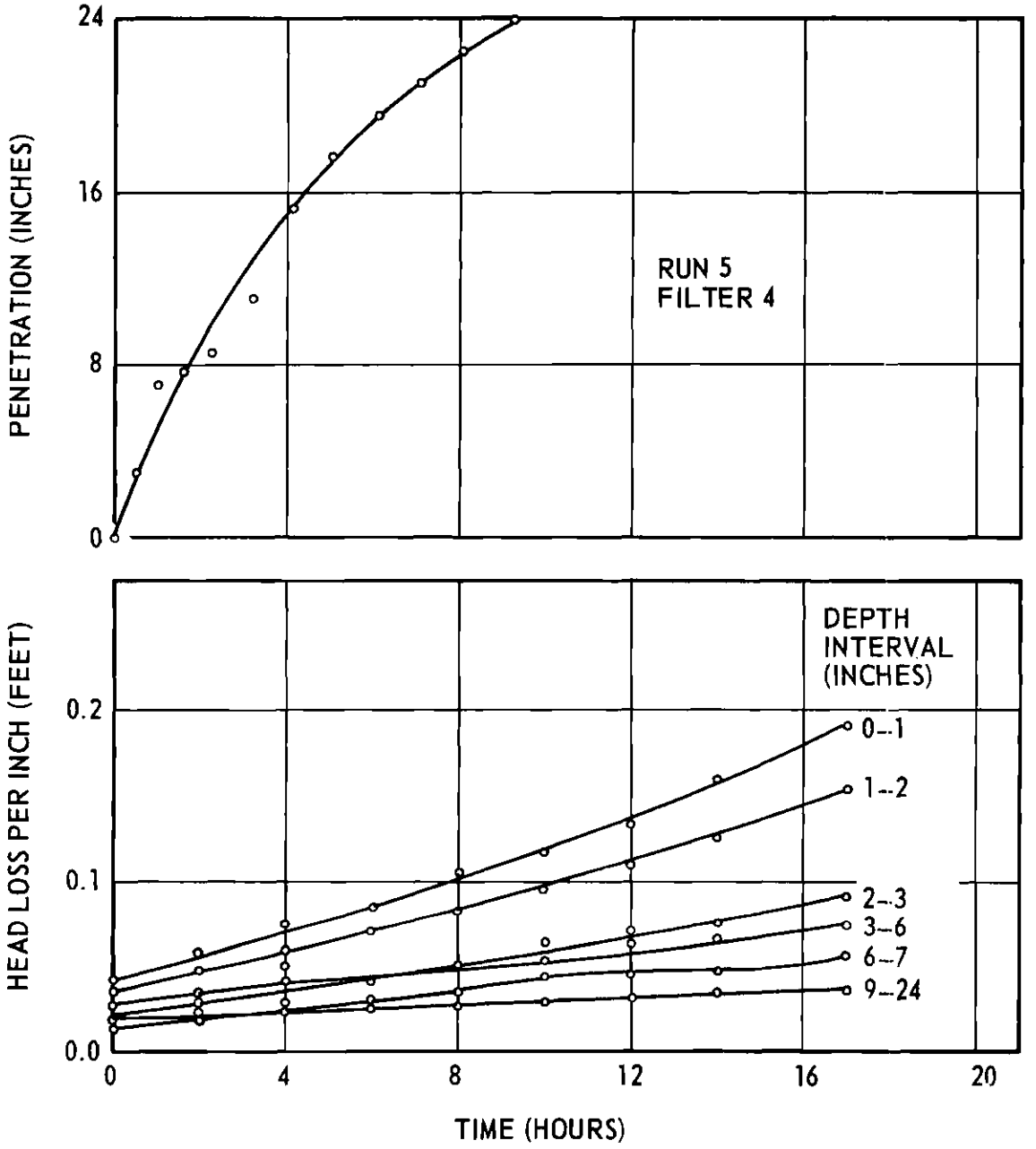


Figure 23. Bed Penetration and Head Loss per Inch for Filter with Removal Distributed Throughout Filter Bed.

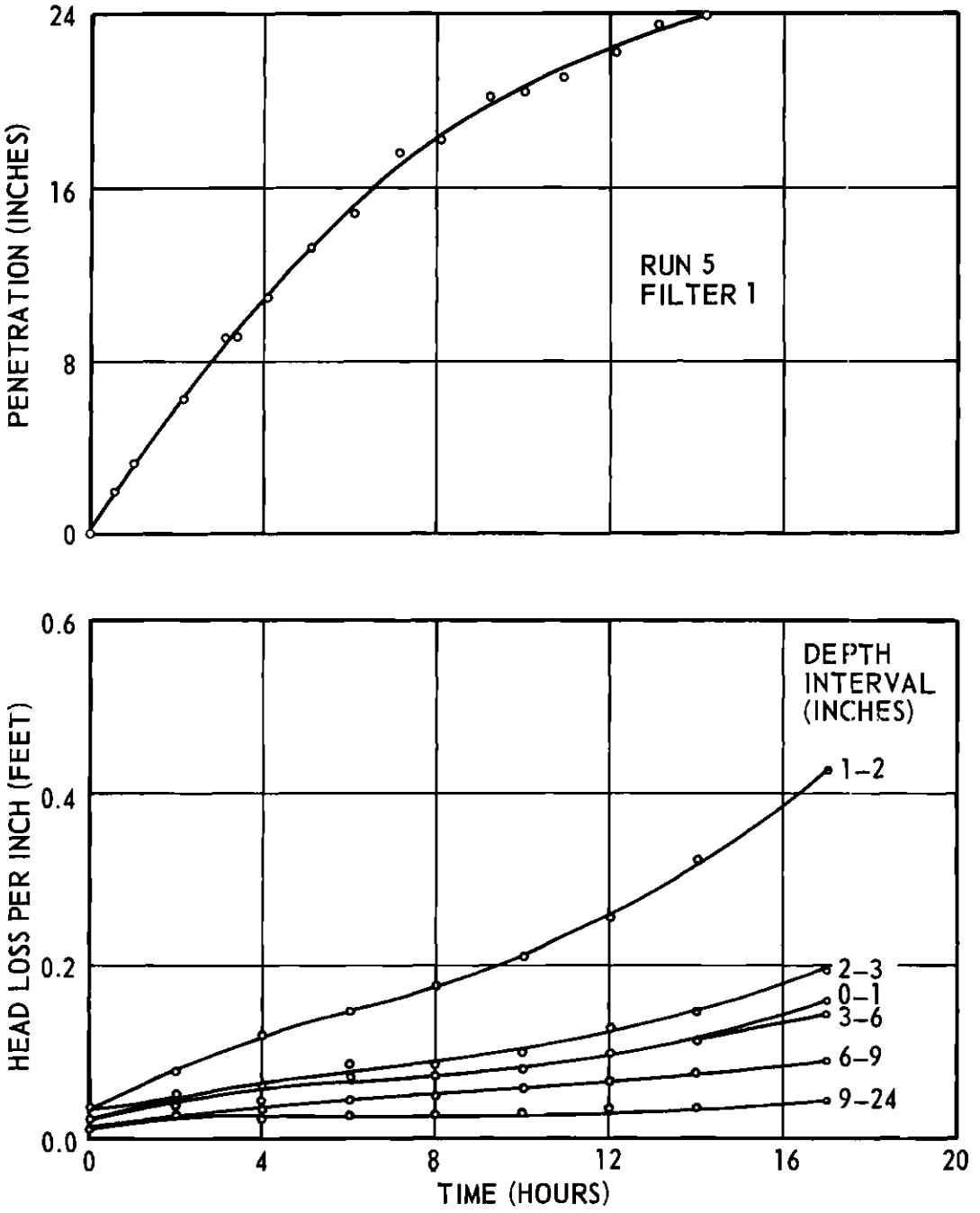


Figure 24. Bed Penetration and Head Loss per Inch for Filter with Greater Head Loss at Intervals Within the Filter Bed than at the Surface.

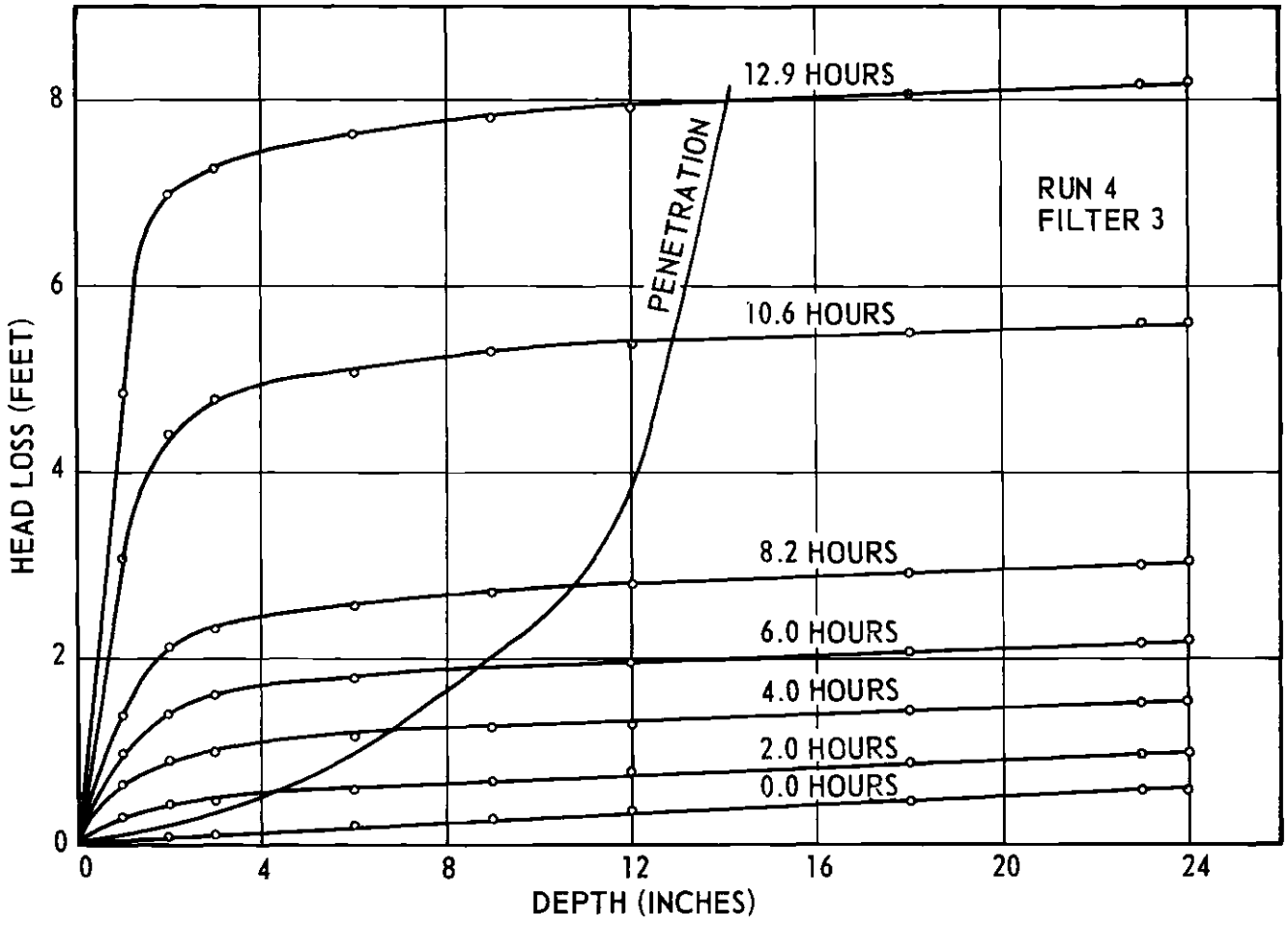


Figure 25. Head Loss Versus Bed Depth for Several Filteration Times.

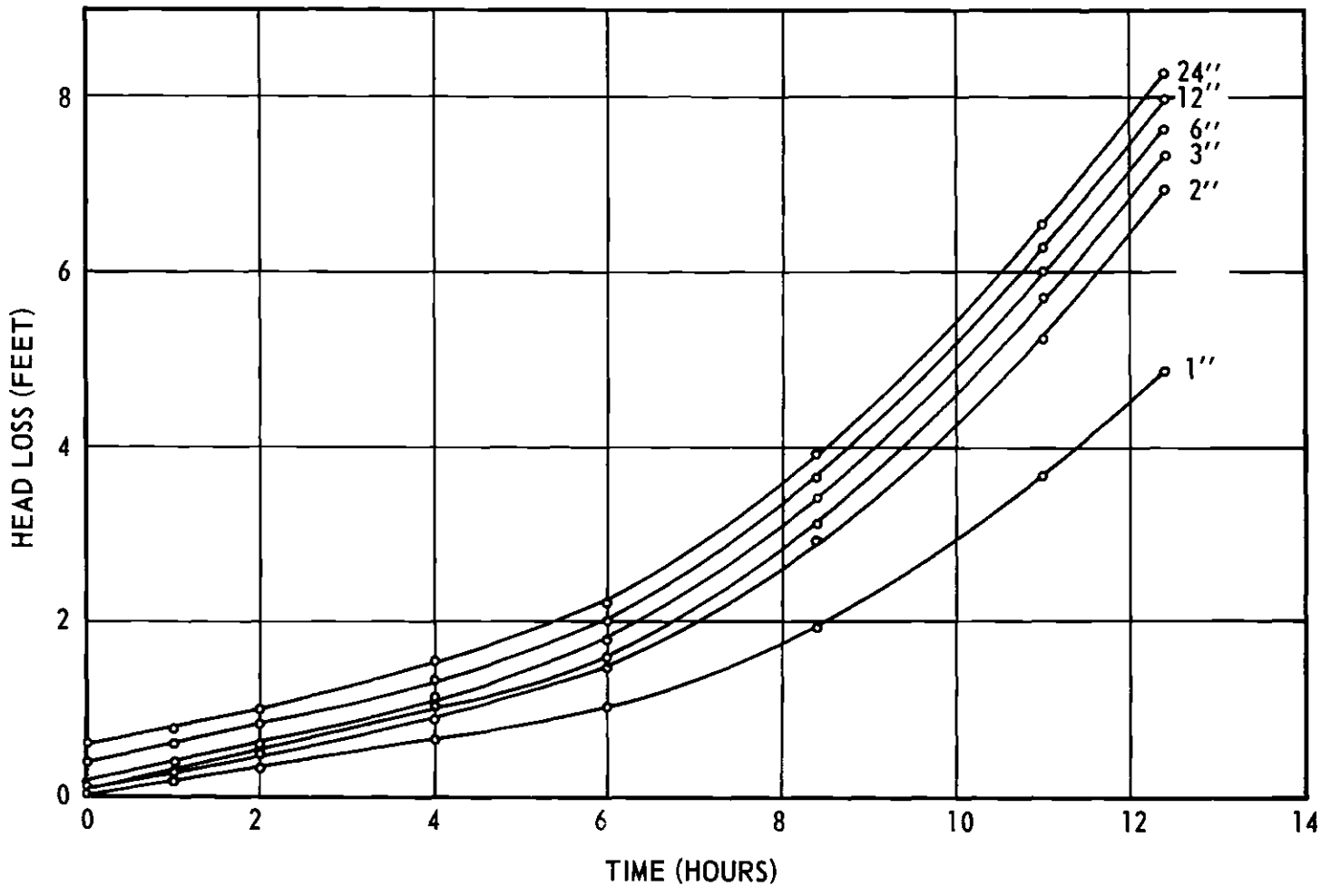


Figure 26. Head Loss Versus Filteration Time for Several Bed Depths.

disturbance resulted in a decrease in head loss but the system recovered in about a half hour. In order to present the results in the manner that would have occurred without the mishap, all head loss data in Run 4 after 6.00 hours of elapsed time were shifted one half hour toward zero time. A similar mishap occurred in Run 5 for the filter receiving the sulfate system. All data for this filter after 9.1 hours of operation have been shifted 1.5 hours toward the origin. The actual time and adjusted times are given in Appendix A for the interested reader. These mishaps did not seriously affect bed penetration and these data were not adjusted.

Since the data for Runs 4, 5, and 6 are compared not only between filters during each run but also between filters during different runs, the head loss and penetration data were plotted against total iron applied. This eliminates any variation resulting from differences in applied iron concentration. The abscissa (mg of ferric iron applied) is computed from the product of the flow rate (2 gpm/sq ft), the filtration time, and the average applied iron concentration for each filter.

It was felt that a more complete understanding of the chemical aspects of sand filtration would require some knowledge of the effects of the chemical composition of the aqueous phase on the surface of the filter medium. Table 5 summarizes the results of experiments conducted on sand with the same aqueous solutions used in the filter runs. The effects of pH on the electrophoretic mobility of the sand are shown in Figure 27.

All data used in preparation of the figures are compiled in the Appendices.

Table 5. Electrophoretic Mobility of Sand Allowed to Reach Equilibrium with Aqueous Solutions Used in Filtration Research

	Filter 1	Filter 2	Filter 3	Filter 4
Run 3				
Mobility	-3.20	-1.417	-2.81	-2.81
pH	10.45	6.02	5.98	6.00
Resistance	8,700	8,000	11,600	9,100
Added salt	Na_3PO_4	$MgCl_2$	Na_2SO_4	NaCl
Run 4				
Mobility	-2.10	-2.38	-2.76	-2.05
pH	5.16	5.12	5.12	5.00
Resistance	296,000	11,400	14,800	41,000
Added salt	None	NaCl	Na_2SO_4	NaH_2PO_4
Run 5				
Mobility	-2.22	-2.22	-2.35	-2.54
pH	8.10	8.63	8.12	8.65
Resistance	20,000	7,500	9,000	9,900
Added salt	None	NaCl	Na_2SO_4	Na_2HPO_4
Run 6				
Mobility	-2.35	-2.70	-2.61	-2.46
pH	6.90	6.78	6.72	7.10
Resistance	110,000	10,500	13,100	24,600
Added salt	None	NaCl	Na_2SO_4	NaH_2PO_4

Note: See Tables 1, 2, 3 and 4 for more information on the aqueous systems used.

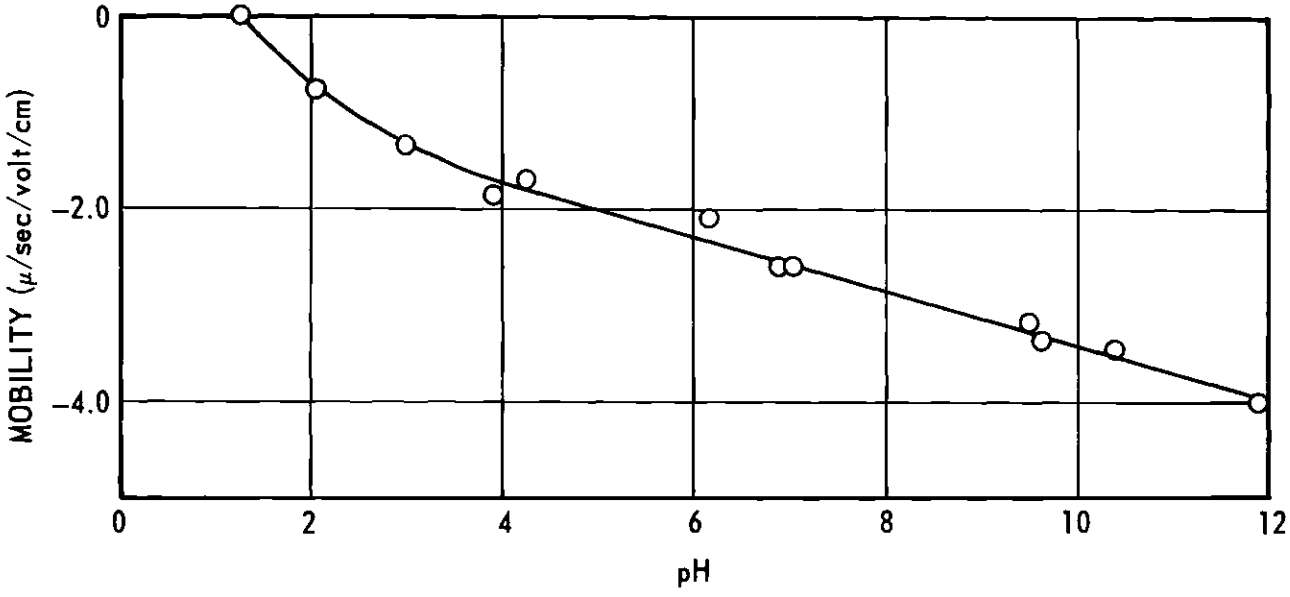


Figure 27. The Effect of pH on the Electrophoretic Mobility of Sand.

CHAPTER VII

DISCUSSION OF RESULTS

The first objective of this research was to evaluate the importance of chemical aspects in the sand filtration process. The second objective was to determine the effects of selected chemical systems on the parameters used in sand filter design. The third aim was to determine the principal particle collection mechanism in the filtration of flocculant suspensions. The following discussion presents the results of this research in light of these goals.

A total of six filter runs were made utilizing four filters during each run. The results obtained from an investigation of sixteen chemical systems are analyzed in this discussion.

An inspection of the data presented in Figure 12 reveals that significant differences in filtration can be produced by a change in the chemical composition of the suspending medium. It was planned to study the effects of 50 mg/l of NaCl, Na_2SO_4, Na_3PO_4, and $MgCl_2$. This combination provided cations and anions of different valence for study. All were at approximately the same pH (6.5) except the Na_3PO_4 system which was at a pH of 10.2. The phosphate pump was inoperative except for one hour period (shown in Figure 12) and the resulting pH was close to that of the other filters for most of the run. There was little increase in head loss during this one hour period, after which the rate of head loss started increasing at approximately the same rate as before the phosphate

was added to the filter. Unfortunately, no penetration data were taken during this period.

In order to gain a better understanding of the filtration process Runs 4, 5, and 6 were designed. It was decided to study the effects of selected anions on the filtration process with one of the four filters being used as a control. Stumm and Morgan[61] discussed the effects of chloride, sulfate, and phosphate ions on coagulation with ferric salts. Since some information was available concerning the effects of these anions on the floc they were selected for study. Twenty-five mg/l of each anion were employed at pH levels of 5.0, 7.0, and 9.5. This concentration and range of pH is within the limits encountered in natural waters[96] in this country and provided a wide range in floc surface characteristics.

An investigation of the data presented in Figures 13 through 19 reveals that wide differences in removal, as characterized by bed penetration and head loss, were demonstrated. Since normal parameters such as sand size (-20 + 30 mesh), flow rate (2 gpm/sq ft), bed depth (24 inches), and bed porosity (40 per cent) were maintained at constant levels, the observed differences are attributed to the chemical composition of the aqueous solutions. The data of Figures 13 through 15 compare the effects of each aqueous system on the filtration process at a constant pH while the data presented in Figures 16 through 19 compare the effect of pH on each chemical system. At constant pH phosphate ions produced the greatest bed penetration and least head loss, generally followed by the sulfate and then the chloride systems. With each aqueous system the poorest filtration occurred at a pH level of 9.5.

Comparison of the filtration characteristics depicted in Figures 13,

14, and 15 with the floc mobilities shown in Tables 2, 3, and 4 indicates that there is a trend towards more difficult filtration as the electrophoretic mobility of the applied floc particles becomes increasingly negative. This trend is summarized qualitatively in Table 6, which compares the relative ease of particle removal in these aqueous systems as indicated by bed penetration, head loss, and electrophoretic mobility of the applied floc particles. The data presented in Table 7 compares the ease of removal for each aqueous system at various pH values.

These qualitative comparisons are fair, but attempts to quantitatively correlate the electrophoretic mobility (or zeta potential) of the suspended floc particles with head loss and bed penetration have been unsuccessful. It was originally proposed that electrokinetic effects in filtration would be due to interactions between the suspended floc particles themselves. Analysis of the data obtained in this research leads to the conclusion that chemical effects are quite significant but are more easily interpreted as an interaction between the surfaces of the filter medium and of the suspended floc particles. Sand exhibits a negative surface charge when placed in contact with water and would, therefore, tend to repel those particles which exhibit a negative electrophoretic mobility while adsorbing those materials which have a positive surface charge. Particles close to the isoelectric point would also adhere to the sand surface if the forces of attraction are greater than the repulsive forces.

In order to ascertain the effects of the different ions on the surface properties of the sand, floated silica (about 240 mesh) was obtained for study. It was felt that this fine sand would define the rel-

Table 6. Relative Ease of Particle Removal with Various Chemical Systems

pH	Indicator	Relative Ease of Removal
5.0	bed penetration	phosphate < sulfate << chloride < demineralized water
	head loss	phosphate << sulfate << demineralized water < chloride
	electrophoretic mobility	phosphate << sulfate << demineralized water ≅ chloride
7.0	bed penetration	phosphate <<< sulfate < chloride < demineralized water
	head loss	phosphate << sulfate << demineralized water << chloride
	electrophoretic mobility	phosphate << sulfate ≅ demineralized water ≅ chloride
9.5	bed penetration	phosphate < demineralized water < chloride < sulfate
	head loss	phosphate ≅ demineralized water << chloride < sulfate
	electrophoretic mobility	phosphate < demineralized water < sulfate < chloride

Table 7. Relative Ease of Particle Removal for Each Aqueous Systems at Various pH Levels

System	Indicator	Relative Ease of Removal
Demineralized Water	bed penetration	pH 9.5 <<< pH 7.0 < pH 5.0
	head loss	pH 9.5 <<< pH 7.0 < pH 5.0
	electrophoretic mobility	pH 9.5 << pH 7.0 << pH 5.0
Chloride	bed penetration	pH 9.5 << pH 7.0 < pH 5.0
	head loss	pH 9.5 <<< pH 5.0 < pH 7.0
	electrophoretic mobility	pH 9.5 < pH 7.0 << pH 5.0
Sulfate	bed penetration	pH 9.5 << pH 5.0 < pH 7.0
	head loss	pH 9.5 << pH 5.0 << pH 7.0
	electrophoretic mobility	pH 9.5 << pH 7.0 << pH 5.0
Phosphate	bed penetration	pH 9.5 < pH 7.0 << pH 5.0
	head loss	pH 9.5 << pH 5.0 < pH 7.0
	electrophoretic mobility	pH 9.5 < pH 7.0 < pH 5.0

ative effects of various aqueous systems on the filter media. The filter sand itself was too large to determine its electrophoretic mobility using available equipment. A mixture of silica and distilled water was prepared at a convenient concentration for mobility measurements. Mobility values were determined, indicating highly negative surface charges in the distilled water. A check of this suspension on the following day gave a different value, lower than the first. It soon became evident that time is required for equilibrium to be attained.

A review of the literature revealed extensive previous work on silica surfaces in aqueous systems. Wood[97] observed that some time elapsed before silica assumed an equilibrium potential in water, which he attributed to effects amounting to a hydration of the surface. He also noted a gradual decrease in potential with time. Wood utilized two cells for study. When the two cells reached identical values, this was taken as the most probable value for the equilibrium potential. A probable value of -177 millivolts is reported for the zeta potential of vitreous silica in pure water. This value was determined by a streaming potential method.

Jones and Wood[98] measured the zeta potential at the interface between vitreous silica and solutions of potassium chloride by the streaming potential method. Their results indicate a decrease toward zero potential with an increase in potassium chloride concentration. Measurements with 10^{-3}, 10^{-4}, and 10^{-5} normal KCl solutions gave zeta potentials of -112.5, -133.6, and -142 millivolts, respectively.

Wood and Robinson[99] investigated the effects of barium chloride on the zeta potential of vitreous silica. The zeta potential was found

to decrease in a regular manner with increasing concentration. The barium chloride decreased the potential more than the potassium chloride at the same concentration.

Wood[100] found that the addition of a trace of thorium chloride to potassium chloride solutions reversed the zeta potential at the silica interface. Vitreous silica, which was strongly negative in contact with the KCl solutions without added throium, exhibited positive values of smaller magnitude.

Wood and Robinson[101] investigated the effect of thorium chloride and lanthanum chloride on the zeta potential of vitreous silica. In general, thorium chloride had a greater effect than lanthanum chloride at the same concentration. Figure 28[101] summarizes the work of Wood et al. with vitreous silica in contact with solutions of potassium chloride, barium chloride, lanthanum chloride, thorium chloride, and distilled water.

Elton et al.[102-107] verified the work of Wood using the sedimentation method and extended it to include the effects of different anions. Dilute aqueous solutions of sulfuric acid, potassium sulfate, potassium ferrocyanide, and potassium ferricyanide were used. Their results indicate that the valence of the anion has little effect on the magnitude of the charge and potential for a negatively charged surface, the important factor being the valence of the cation.

Table 5 is a summary of the equilibrium electrophoretic mobilities exhibited by sand particles placed in the aqueous systems used in Runs 3 through 6. In general, there is little difference between the values of the mobility for Runs 4 through 6. During Run 3 wider differences in

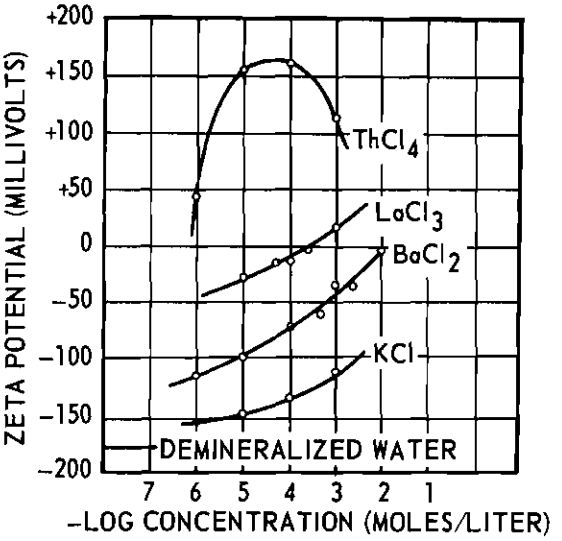


Figure 28. Surface Characteristics of Vitreous Silica (After Wood and Robinson (101)).

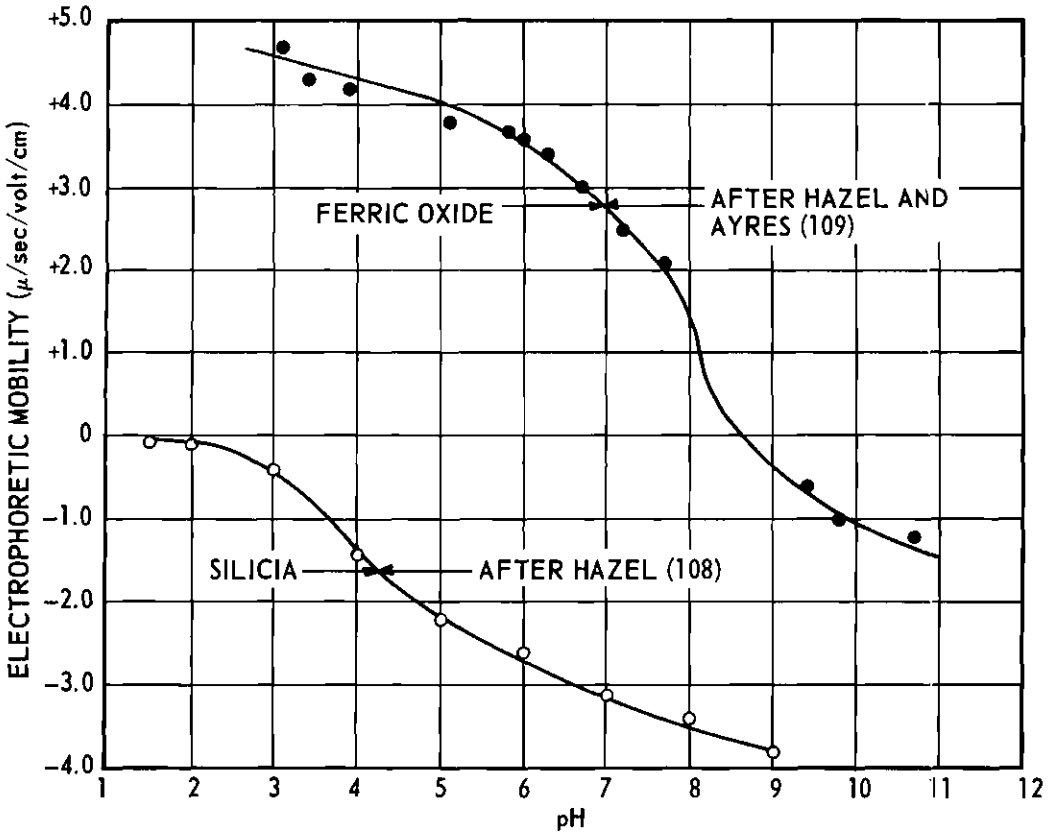


Figure 29. The Effects of pH on the Electrophoretic Mobility of Silica and Ferric Oxide (After Hazel (108) and Hazel and Ayres (109)).

sand mobility are noted. The $MgCl_2$ reduced the mobility of the sand more than any of the other chemical systems. This agrees with the work of Elton et al.[102-107], since the divalent cation produced a greater effect than the monovalent cations in the other systems. This also confirms that the cation is more important than the anion in determining the zeta potential of the sand.

The effects of pH on the electrophoretic mobility of the silica sand are presented in Figure 27. Considerable difficulty was encountered in obtaining equilibrium values of the mobility of the sand. The studies summarized in Table 5 and Figure 27 were obtained by first placing the sand in distilled water and letting it hydrate for two weeks. Samples of the sand were then placed in the aqueous systems to be studied and allowed to stand for a third week. Numerous trials were made before equilibrium values could be obtained but the procedure outlined yielded good results. Figure 27 compares very favorably with the work of Hazel[108] for silica as shown in Figure 29. Superimposed on Figure 29 is the effect of pH on the mobility of ferric oxide as determined by Hazel and Ayres[109] for comparison with the curve obtained for silica.

The knowledge gained about the silica surfaces in the laboratory and in the literature provided a better background upon which to postulate the removal mechanism for the aqueous floc suspensions studied. In all aqueous systems studied the poorest filtration as characterized by low head loss increase and rapid bed penetration occurred at a pH of 9.5. These effects may be explained by the fact that both floc and sand surfaces exhibited negative mobilities and consequently negative zeta potentials. An additianal aspect to be considered is the fact that the ionic

strength of the aqueous systems at pH 9.5 was considerably higher than at other pH levels, due to the bicarbonate-carbonate buffer system employed. This high ionic strength would produce a reduced double layer thickness and an increased charge density for the materials under investigation.

The phosphate ions produced the greatest effect on filtration of the systems studied. The electrophoretic mobility of these floc particles were more negative at each pH than were those for the other systems. The effects observed due to the phosphate ions cannot be explained solely on the basis of the valence of the anion involved. Calculations based on the appropriate ionization constants show that a pH 5.0 essentially all of the dissolved phosphate was present as monovalent $H_2PO_4^{-1}$ ions, at pH 9.5 it was essentially all divalent HPO_4^{-2} ions, and at pH 7.0 it was divided into approximately 38 per cent divalent (HPO_4^{-2}) and 62 per cent monovalent ($H_2PO_4^{-1}$) ions. Experiments performed with divalent sulfate ions at each of these pH levels produced better floc removal than with the phosphate systems. This is not what would be expected considering valence only.

Additional insight into the effects of phosphate ions on the filtration process can be obtained from an inspection of the data presented in Figures 12, 20, and 21. During Run 3 the phosphate pump operated for only a one hour period with a very significant decrease in rate of head loss increase for this hour. Unfortunately no penetration data were taken during this period but most probably the penetration increased significantly during this time. The electrophoretic mobility of the floc and sand were highly negative during this one hour period during which 50 mg/l

of Na_3PO_4 were applied with a resulting pH of 10.2. The mobility of the floc increased significantly (from -1.36 to -1.99 μ/sec/volt/cm) during this period.

During Run 4, the phosphate pump was shut off after 7.25 hours of operation. The phosphate concentration was therefore reduced to almost zero with only that originating from the stock floc suspension being present. Almost immediately significant changes occurred in both the head loss and bed penetration as shown in Figure 20. The rate of change of head loss with time increased sharply, and approximated the rate observed for a demineralized water system at this pH. Bed penetration was halted almost completely for about 2.5 hours, after which it increased at a rate somewhat similar to that observed for a demineralized water system. O'Melia[77] has shown that an increase in phosphate ion concentration increases the penetration of iron into the bed which confirms the results of this experiment.

It is postulated that the floc removed during the first 7.25 hours of this run coated the sand grains and produced a negatively charged interface with the aqueous phase. When the dissolved phosphate was removed from solution, the surfaces of both the suspended floc particles and those which had previously been retained in the bed were altered, and their surface charge decreased towards the isoelectric point. Unfortuantely the mobility samples taken during this period of the run were badly contaminated by acid adsorbed onto the sample bottles. Later O'Melia[77] investigated the effects of placing floc prepared in a phosphate environment into demineralized water at pH 5.0. The mobility of the floc decreased from a negative value to a value close to the isoelectric point. This change

in surface characteristics of both the applied particles and the upper strata of the filter medium enabled iron floc to be applied to the filter for an additional 2.5 hours without further penetration into the bed. After this time the applied materials resumed their penetration into the bed at a reduced rate. It is felt that this observation is strong support for the concept that the removal of floc particles in a conventional sand filter is accomplished by adsorption to the filter medium and that this adsorption is materially affected by the chemical composition of the aqueous phase.

The data of Figure 21 gives support to the adsorption theory. These data represent the head loss for several depth intervals at different times. Until 7.25 hours the removal as characterized by the distribution of head loss through the filter bed was fairly evenly distributed. After removal of the phosphate pump from service, the floc were removed primarily in the upper strata of the bed as shown by the very high distribution of head loss for this region. Bed penetration and head loss per inch for a filter with all the removal near the surface is shown in Figure 22. The distribution of head loss in Figure 21 after 7.25 hours resembles the distribution shown in this figure.

Bed penetration and head loss per inch for a filter with removal distributed throughout the filter bed is shown in Figure 23. Again the comparison is made between Figure 21 and this figure. Until 7.25 hours of filtration the comparison is good.

Review of Figures 22 and 23 should illustrate the danger of advocating shallow filter depths for filters unless more is known about the effects that the ions in solution in the water in question have on

the filtration process. Many engineers believe that the 24 inches of sand used in most filters provides a great safety factor. If the filtration system performs as the one shown in Figure 22 this may be true. If it performs as the one shown in Figure 23 there is no safety factor if operation is based on terminating the run after all the available head loss is utilized.

The bed penetration and head loss per inch for a filter with more head loss at intervals within the filter bed than at the surface is shown in Figure 24. Several explanations can be offered. One is that the top bed interval was not actually an inch due to improper preparation of the sand bed. The writer believes that this is not the case since a significant difference from the other filters would have been noted during the hourly bed penetration measurements. A second explanation is that a channel was opened from the surface to the one inch manometer point. This is possible. Another possible explanation would be coagulation within the filter bed. The zeta potential of the floc for this run had a mobility of -1.72 μ/sec/volt/cm which makes this unlikely. However, the high ionic concentration present in Run 5 would result in a very thin diffuse layer which could allow close contact of the floc before significant repulsion resulted. If the van der Waals forces of attraction were greater than the repulsive forces coagulation could occur.

Figure 25 represents the relationship observed between head loss and bed depth at various times for Filter 3 during Run 4. It can be noted that at zero time the head loss varies directly with the bed depth in a linear manner, a relationship which has been observed by several investigators in the past. Superimposed upon this curve is the bed pene-

tration curve for this filter. In general the head loss at each depth varies linearly until the iron has penetrated to that depth at which time this linear relation disappears.

Figure 26 depicts the observed head loss versus filtration time for several bed depths. This figure presents the data given in Figure 25 differently. O'Melia[4] observed a linear distribution of head loss regardless of the depth of bed considered in the filtration of algal suspensions. The relationship is linear for only a short time in Figure 26 at which time the head loss at all depths considered increases at an exponential rate.

The curves are generally parallel except for the intervals near the surface. The rate of clogging is worst near the surface which explains a deviation from the near parallel relationship observed between other depths.

The experimental results give strong evidence of an interaction between two oxides (SiO_2 and Fe_2O_3) with different degrees of hydration and different ions incorporated into their electrical double layers. Some general information on the effects of ions on silica surfaces has already been presented. It is believed that a review of the effects of ions on the ferric system would also be helpful.

Hazel and Ayres[109] studied the effects of different aqueous systems on positive ferric oxide sols. Their results indicate that the valence of the anion generally determines the relative effects of the ions. Their results are summarized in Figure 30[109]. It was found that by the addition of potassium ferrocyanide the charge of the ferric oxide sols could be made negative. In a later study[110] they investigated the effects

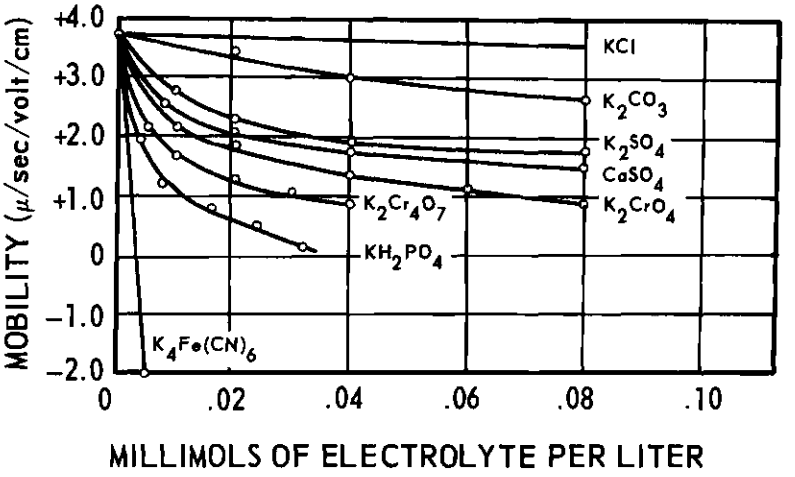


Figure 30. Surface Characteristics of Ferric Oxide Sols (After Hazel and Ayres (109)).

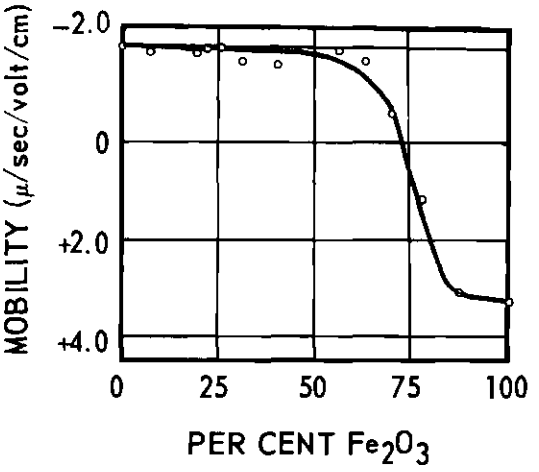


Figure 31. Mobilities of Silica Ferric Oxide Mixtures (After Hazel (108)).

of different aqueous solutions on negative sols prepared by addition of ferrocyanide. Potassium chloride, barium chloride, aluminum chloride, and thorium nitrate were studied. The greater the valence of the cation, the greater the effect at the same concentration. Aluminum chloride and thorium nitrate in higher concentrations made the sols positive with the thorium having the greater effect. In general the anion has a greater effect on positive sols while the cation has the greater effect on negative sols.

In some preliminary runs made by Stanley[32] to determine a suitable material for filtration studies, jeweler's rouge (Fe_2O_3) was employed. He reported 100 per cent removal of the rouge by the sand filter but also reported that no amount of backwashing would remove it. An investigation of Figure 29 reveals that ferric oxide would be strongly attracted by the silica at pH levels below about 8.6 since the ferric oxide is positive and the silica negative. The pH level at which the study was made was not stated but was probably 7.0 or less since Stanley used demineralized water in his filtration studies.

Electrokinetic forces are the most important particle collection mechanism in sand filtration in the opinion of the writer. The interaction between the electrokinetic forces of the sand and the suspended particles determine the filtrability of aqueous floc suspensions. Since the chemical composition of the aqueous phase affects the electrokinetic properties of both the sand and the suspended particles and since there is an interaction between suspended matter and the sand surface, the entire system must be considered as a whole. For flocculant materials such as the hydrated oxides of iron and aluminum, the surface charge of the sand

is changed by the presence of these materials. If salts are added, changes are brought about in both the floc and sand surfaces which can significantly alter the effects that the floc and sand have on each other. Since interactions of these electrokinetic forces are masked by a buildup of head loss in the filter, with possible shear and further penetration into the filter bed, the writer proposes that prediction of filter performance could better be made by electrophoretic studies similar to those conducted by Hazel[108] in his study of the mutual coagulation of silica and ferric oxides. If the effects of added ions were determined for a ferric oxide system and then for a silica system, the two systems could be combined and the net effects studied.

Hazel[108] studied a system of mixed colloidal silica and ferric oxide. He found that more than 70 per cent ferric oxide was required to produce an isoelectric mixture. Over 60 per cent ferric oxide was required before an appreciable drop in mobility of the silica was noted. This indicates a great capacity for adsorption of floc particles. Hazel also studied the mutual coagulation of aluminum oxide and silica. He found aluminum oxide to be much more effective in discharging silica than ferric oxide, requiring only 17 per cent for an isoelectric mixture. These studies were conducted at a pH of 5.1 utilizing well dialyzed solutions. Figure 31 summarizes Hazel's results[108] for ferric oxide-silica mixtures.

Hazel[108] also studied the effects of sodium chloride, potassium chloride, calcium chloride, barium chloride, and aluminum chloride on the mobility of colloidal silica. In general, cations of the same valence produced similar results with the higher the valence the greater the effect.

Increasing concentrations of all salts decreased the magnitude of the mobility (from negative toward zero). These results support the findings of others already presented.

O'Connor and Buchanan[111] indicate that the effects of electrolytes on the surfaces of insoluble oxides are best interpreted in terms of both preferential ion adsorption (physical) and ion exchange involving the hydroxyl groups of the hydrated surface layers. The relative importance of the anion and cation would then depend on the hydrogen ion concentration of the solution. They state that it is very probable that the surfaces of insoluble oxides undergo reaction with water to adsorb hydroxyl groups, producing a surface which may ionize either as a weak acid or a weak base, depending on the circumstances. The electrical double layer established in this way will have either hydrogen or hydroxyl ions in the diffuse layer when the substances are in contact with pure water. There is considerable evidence to indicate that silica binds water to give a surface layer of silicic acid. Since silica has a negative charge in pure water the surface hydroxyl groups are behaving as acids.

The persistent negative charge displayed by all forms of silica suggests that the surface hydroxyl groups mainly responsible for the charge are comparatively strongly polarized by the coordinating silicon atoms, thus facilitating the release of hydrogen ions when in contact with water. Therefore there is little tendency for silicic acid to exhibit basic properties analogous to those of the hydrated oxides of other metals.

Gaudin et al.[112] have used tracer techniques in experiments on reaction of sodium salts with quartz in which good evidence for the exis-

tence of cation exchange has been presented. Ion exchange (both cation and anion) has been offered as the explanation of changes in surface charge of oxides. Ions can be physically adsorbed by a surface already possessing a charge of the same sign. The extent of adsorption is independent of reasonable changes in the magnitude of the charge on the surface[113]. For this reason surfaces with a negative charge can be made more negative by addition of certain anions since the surface may be preferential to that type of ion.

Johansen and Buchanan[113] report that potentials could be returned to their normal value (that in water) by prolonged washing as long as the ions were of the same sign. Cation adsorption by negative surfaces could not be reversed by washing. This is probably the case in a sand filter. Upon addition of coagulated materials, the dissolved iron (or aluminum) left in solution would be adsorbed by the sand surface thereby reducing the potential. Other ions in solution may also be preferentially adsorbed. During backwashing of the filter, the sand probably regains part of its original charge but the reduction in charge due to cation adsorption would not be reversed completely, if at all. The addition of chlorine and other chemicals added after filtration would be present in the backwash water and would affect the degree of restoration of the charge.

During Run 2 it was pointed out that deeper penetration was observed for the new sand than for the old sand. The electrophoretic mobility of the floc was -1.22 μ/sec/volt/cm and it is safe to assume that the charge on the sand was reduced by the iron adsorbed during Run 1. With a reduced surface charge on the sand, less repulsion would be expected and better

removal achieved, as observed.

The writer believes that for flocculant materials the following events can be considered: (1) If the floc are positive and the sand negative, the floc will be adsorbed onto the sand surface thereby decreasing the diameter of the pores in the filter bed. Additional removal beyond the adsorption capacity of the sand would be due to straining. (2) If the floc have a charge close to zero or in general if the attractive forces of the floc exceed the repulsive forces, one of two things probably occurs. The first possibility is coagulation of the particles within the bed. Coagulation would result in an increase in the average diameter of the floc and removal would be effected by straining. The second possibility is the adhesion of the floc to the sand surface which will be called adsorption as in case one. Mechanisms are available for transport of particles to the sand surface and with little repulsion due to a low zeta potential, the floc would adhere to the sand surface or to floc previously removed.

The removal of "inert" suspended matter by a filter is pictured somewhat differently but still depending on the electrokinetic forces present. Clay particles, algae, bacteria, and viruses, to mention a few, are removed by filters to a degree even though the particles are much smaller than the pore diameter of the filter. Hunter and Alexander[84] have proposed a diffusion of these small particles into "dead spaces" in the filter due to a concentration gradient. Electrokinetic forces of attraction between the filter medium and the suspended materials could increase the probability of a particle diffusing into one of these areas of low shear. In one experiment the sand was rendered positive by the

adsorption of hexadecyltrimethylammonium ions. It was reported that the general features of the deposition process were unchanged though the amount of deposition was substantially increased. It is possible that in this case the attractive forces between the sand and clay particles increased the probability of diffusion of the particles into the dead spaces. Since the clay particles do not have the adhesion characteristics exhibited by flocculant materials, the clay particles made contact with the sand but were probably removed after contact by hydrodynamic shear forces except in the dead spaces.

An increase in the repulsive forces between the suspended particles could also increase the probability of a particle diffusing into the dead spaces due to mutual repulsion of the suspended particles. Jordan[83] reported improved filtration with increasing negative zeta potential, and offered as a possible explanation the presence of a positively charged Schmutzdecke. Jordan agrees that a positively charged Schmutzdecke is hard to conceive of since it is composed of bacteria, clay, and other materials generally considered to be negatively charged, but offers the possibility that the products of decomposition of these materials may be positive. It is also possible that the particles were removed by the mechanism proposed by Hunter and Alexander[84] and that the increasing negative potential of the suspensions studied contributed to the particle transport by diffusion and therefore produced better removal. It is also noted that investigation of Jordan's work indicates that improved removal efficiencies correspond with elapsed time of filtration. This could correspond to the period of "ripening" noted by other investigators.

The removal of viruses, bacteria, and algae which are negatively

charged is enhanced by the addition of coagulants. The coagulation of these materials would be similar to coagulation of clay and other colloidal particles of negative charge. It is believed that the addition of positively charged coagulants reverses the net surface charge of these suspended particles. In turn the positively coated materials are adsorbed onto the sand surface.

The work presented in this thesis disagrees somewhat with the work of Stanley[31,32]. Stanley reported very poor filtration at low pH levels and best filtration at a pH near 7.0 In most cases the writer observed best filtration at a pH of 5.0 with reduction in pH accompanied by decreasing negative values of floc mobility. It should be pointed out that Stanley employed high salt concentrations in preparation of his radioactive floc. In addition, the negative iodide ions were absorbed into the floc. The characteristics of both the floc and sand were probably changed by the high ionic concentration and possibly even by the radiation. Additional information may be gained from his investigation of the effects of 500 mg/l of NaCl, Na_2SO_4, and $MgSO_4$ on filtration. He reported 65 to 70 per cent increase in "Penetration Index" for Na_2SO_4 and $MgSO_4$ and about 15% for NaCl. This indicates to the writer that the double layer thickness was already thin before adding these salts. Based on knowledge gained during this research a wider difference would probably be expected with "normal" floc. After demonstrating these differences in filtration due to pH and the presence of selected chemicals Stanley developed an empirical equation for penetration involving only flow rate and sand size, indicating that he did not consider the results of his chemical studies as significant in conventional water treatment

systems.

Sanford and Gates[75] reported that positive sand did not remove either bacteria or aluminum hydroxide better than normal sand. It should be noted that no measurements were made to determine the surface potentials of either the sand or the suspended matter. They concluded that adsorption is not of importance in filtration. It should be noted that this conclusion was based on assumptions made about the charge of the sand and suspended matter.

These authors did report increased head loss for the filters in which adsorption could have occurred. No differences in depth of penetration were noted. A possible explanation for this is that the floc initially were removed near the surface by adsorption and penetrated deeper after shear increased in the upper layers. The writer is convinced that floc do move deeper into the bed after being removed in a filter. This is based on the high (as much as five times the influent concentration) concentration of iron in samples carefully taken from within the filter bed during Runs 1 through 3. This would of course depend on how strongly the floc are bound to the bed and/or other floc particles.

The results presented in this thesis have practical applications in filtration. A wide range in bed penetration and head loss can occur with changes in the chemical composition of the aqueous suspending medium. A change in the chemical composition of the water is more easily effected than changes in sand size, bed depth, or flow rate after the filter has been constructed. Optimum filtration can then be controlled by addition of appropriate chemicals.

The ideal filter would share the removal burden throughout its

depth, produce water of acceptable quality, and operate for a reasonable length of time resulting in a reasonable loss of head. Addition of phosphate ions result in a decrease in head loss and an increase in bed penetration. Removal of floc ordinarily retained at the surface could therefore be effected at greater depth in the filter bed by addition of phosphate. Elimination of complete bed penetration by other types of suspended matter could possibly be effected by addition of coagulant aids or polyelectrolytes.

It is believed by the writer that filtration will be chemically controlled in the future. More research is needed to determine the chemicals that affect the electrokinetic charge of the sand and suspended medium in a desired manner. With a knowledge of the affects of various chemical systems, complete control of filtration can be effected.

CHAPTER VIII

CONCLUSIONS AND RECOMMENDATIONS

Conclusions

Based upon the experimental results obtained in this research, the following conclusions have been reached.

1. The removal of ferric floc from suspension during rapid sand filtration as evaluated by head loss increase and rate of bed penetration can be significantly affected and, in fact, controlled by the chemical composition of the aqueous suspending medium.
2. Electrokinetic effects resulting from the charged surfaces of the sand filter medium and the suspended floc particles as determined by microelectrophoretic mobility measurements provide the most important particle collection mechanism in the rapid sand filtration of hydrous ferric oxide floc.
3. Phosphate ions exert very significant effects in the filtration of ferric floc, as characterized by a marked decrease in clogging rate and a corresponding increase in the rate of floc penetration into the filter bed.

Recommendations

The following recommendations are offered as a logical continuation of the research presented in this thesis and are directed toward a better understanding of the filtration process.

1. A filtration study should be conducted in which the cation portion of the system is varied. Since the cation has a greater effect on the sand surface than the anion, this may give additional evidence on the importance of the surface charge of the sand.
2. Studies should be conducted in which provisions are made for the measurement of streaming potentials at different depths in the filter. The changes in potential within the filter may then be correlated with head loss and bed penetration.
3. Electrophoretic studies should be conducted in which the effects of ions on a combined system of ferric oxide and silica (as discussed in Chapter VII) be determined. The findings of this study should then be correlated with filtration runs using the ions investigated.
4. A filtration study should be conducted in which provisions are made for measurement of head loss at very small increments throughout the filter bed. A pressure transducer is recommended in order to eliminate passage of flow into and out of the bed as encountered with manometers. More accurate measurements could also be made. A closer look at the changes in pressure within the filter bed would indicate where removal is being effected at all times. It is believed that if the rate of deposition were known at each depth more information could be gained about the mechanism of deposition.
5. A study should be conducted of the adsorption onto and/or displacement of ions from the sand surfaces. A long sand column

could be employed through which demineralized water is passed initially. Conductivity, pH, and samples for chemical analysis should be collected. When the chemical system reaches equilibrium, ions of interest (e.g. chloride, sulfate, phosphate, etc.) should be added. Measurement of conductivity, pH, and ion type and concentration should be made for the filter influent and effluent at regular time intervals and changes noted. Similar techniques could be employed for other chemical systems. Analysis of influent and effluent samples should provide a measure of adsorption and/or displacement of ions. A knowledge of the rate and amount of ion adsorption and/or displacement within the filter bed should be valuable in predicting effects of different aqueous systems on the filtration process.

APPENDICES

APPENDIX A

Head Loss Data

Run No. 3 - Filter No. 1

Bed Depth (inches)	Filtration Time (hours) 0.00	1.00	2.10	3.00	4.00	5.00	6.00
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.025	0.712	1.375	1.271	1.850	2.650	3.250
2	0.050	0.750	1.525	1.529	2.166	3.150	4.017
3	0.075	0.771	1.550	1.562	2.225	3.238	4.171
6	0.150	0.842	1.617	1.629	2.300	3.304	4.238
9	0.208	0.904	1.683	1.696	2.362	3.367	4.296
12	0.275	0.962	1.742	1.754	2.433	3.433	4.354
18	0.379	1.075	1.850	1.866	2.545	3.533	4.467
23	0.462	1.158	1.925	1.954	2.641	3.600	4.550
24 + under-drain	0.479	1.175	1.942	1.970	2.666	3.633	4.567

Bed Depth (inches)	Filtration Time (hours) 7.00	8.00	9.10	10.00	11.00	11.50
Surface*	-	-	-	-	-	-
1	0.000	0.000	0.000	0.000	0.000	0.000
2	1.142	1.275	1.834	1.850	2.321	2.746
3	1.408	1.613	2.217	2.283	2.792	3.246
6	1.475	1.671	2.330	2.417	2.959	3.437
9	1.525	1.742	2.392	2.483	3.017	3.496
12	1.596	1.800	2.446	2.542	3.079	3.546
18	1.708	1.909	2.559	2.650	3.184	3.654
23	1.796	1.992	2.642	2.733	3.267	3.737
24 + under-drain	1.808	2.013	2.659	2.758	3.284	3.754

Note: Phosphate pump operated only from 2.00 to 3.00 hours.

APPENDIX A (Continued)

Run No. 3 - Filter No. 2

Bed Depth (inches)	Filtration Time (hours) 0.00	1.00	2.10	3.00	4.00	5.00	6.00
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.012	0.413	0.808	1.125	1.380	1.712	2.000
2	0.033	0.467	0.958	1.362	1.963	2.442	2.983
3	0.054	0.492	0.983	1.425	2.042	2.579	3.150
6	0.125	0.563	1.050	1.517	2.113	2.662	3.275
9	0.196	0.638	1.117	1.608	2.180	2.725	3.342
12	0.254	0.696	1.175	1.675	2.246	2.787	3.400
18	0.362	0.813	1.292	1.792	2.350	3.896	3.508
23	0.450	0.896	1.367	1.892	2.434	2.979	3.592
24 + under-drain	0.471	0.913	1.392	1.896	2.450	2.996	3.608

Bed Depth (inches)	Filtration Time (hours) 7.00	8.00	9.12	10.00	11.00	11.50
Surface*	-	-	-	-	-	-
1	0.000	0.000	0.000	*-	*-	*-
2	1.313	1.329	1.646	0.000	0.000	0.000
3	1.500	1.475	1.900	0.283	0.325	0.375
6	1.680	1.775	2.200	0.633	0.729	0.783
9	1.746	1.846	2.267	0.700	0.821	0.900
12	1.792	1.892	2.325	0.758	0.883	0.962
18	1.896	2.008	2.425	0.866	0.992	1.075
23	1.988	2.083	2.508	0.941	1.075	1.154
24 + under-drain	2.000	2.096	2.525	0.958	1.092	1.175

APPENDIX A (Continued)

Run No. 3 - Filter No. 3

Bed Depth (inches)	Filtration Time (hours) 0.00	1.00	2.00	3.00	4.00	5.00	6.00
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.029	0.766	1.467	1.888	2.633	3.250	3.600
2	0.059	0.846	1.617	2.358	3.200	4.029	4.934
3	0.092	0.925	1.642	2.388	3.229	4.054	4.983
6	0.167	0.996	1.725	2.467	3.308	4.133	5.042
9	0.242	1.033	1.800	2.542	3.395	4.204	5.117
12	0.309	1.071	1.858	2.609	3.445	4.275	5.188
18	0.434	1.196	1.983	2.738	3.566	4.400	5.309
23	0.525	1.287	2.075	2.834	3.666	4.491	5.409
24 + under-drain	0.546	1.308	2.100	2.854	3.683	4.512	5.434

Bed Depth (inches)	Filtration Time (hours) 7.00	8.00	9.08	10.00	11.00	11.50
Surface*	-	-	-	-	-	-
1	0.000	0.000	0.000	0.000	0.000	0.000
2	1.675	2.196	2.480	3.238	-	3.600
3	1.691	2.234	2.546	3.367	-	3.754
6	1.750	2.309	2.609	3.442	-	3.842
9	1.825	2.367	2.667	3.517	3.841	3.917
12	1.891	2.417	2.738	3.584	3.908	3.984
18	2.016	2.550	2.855	3.709	4.033	4.104
23	2.108	2.642	2.942	3.800	4.125	4.200
24 + under-drain	2.129	2.659	2.963	3.821	4.145	4.221

APPENDIX A (Continued)

Run No. 3 - Filter No. 4

Bed Depth (inches)	Filtration Time (hours) 0.00	1.00	2.00	3.00	4.00	5.00	6.00
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.029	0.562	1.300	1.983	2.521	3.224	3.866
2	0.066	0.608	1.437	2.308	3.113	3.941	4.808
3	0.087	0.633	1.467	2.333	3.134	3.974	4.858
6	0.166	0.716	1.542	2.441	3.209	4.057	4.941
9	0.237	0.791	1.617	2.541	3.279	4.137	5.008
12	0.308	0.866	1.692	2.625	3.350	4.212	5.079
18	0.441	0.991	1.825	2.775	3.475	4.341	5.208
23	0.533	1.091	1.917	2.870	3.559	4.432	5.304
24 + under-drain	0.558	1.112	1.942	2.904	3.575	4.462	5.239

Bed Depth (inches)	Filtration Time (hours) 7.00	8.00	9.07	10.00	11.50
Surface*	-	-	-	-	-
1	0.000	0.000	0.000	0.000	0.000
2	1.508	1.767	2.150	2.504	3.226
3	1.592	1.867	2.279	2.659	3.363
6	1.671	1.954	2.371	2.767	3.496
9	1.742	2.025	2.442	2.842	3.567
12	1.817	2.092	2.517	2.913	3.638
18	1.942	2.175	2.642	3.038	3.763
23	2.037	2.259	2.738	3.142	3.855
24 + under-drain	2.058	2.275	2.758	3.167	3.884

*After 6 hours of operation the head loss exceeded the available manometer length so top manometers were clamped off in order to lower the pressure in the suppressor system. A reference manometer leading up to the constant head tank and a manometer on the suppressor line indicating air pressure in feet of water was installed before Run 4 to eliminate this difficulty. See Figure 10.

APPENDIX A (Continued)

Run No. 4 - Filter No. 1

Bed Depth (inches)	Filtration Time (hours) 0.00	0.50	1.00	1.50	2.00	3.00	4.00
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.029	0.154	0.312	0.496	0.666	1.038	1.425
2	0.062	0.191	0.354	0.563	0.775	1.271	1.829
3	0.091	0.216	0.379	0.592	0.804	1.296	1.858
6	0.166	0.295	0.454	0.667	0.875	1.371	1.929
9	0.233	0.366	0.525	0.738	0.950	1.442	2.000
12	0.304	0.437	0.591	0.800	1.016	1.508	2.066
18	0.429	0.566	0.720	0.929	1.141	1.633	2.196
23	0.524	0.666	0.820	1.025	1.241	1.725	2.292
24 + under-drain	0.550	0.691	0.841	1.083	1.262	1.750	2.312

Bed Depth (inches)	Filtration Time (hours) 5.00	6.00	*6.50 (delete)	7.00 (6.50)	7.75 (7.25)	8.17 (7.67)	8.92 (8.42)
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	1.884	2.333	1.708	2.242	2.675	2.830	3.310
2	2.438	3.112	2.675	3.400	4.025	4.314	4.956
3	2.459	3.166	2.724	3.463	4.100	4.405	5.076
6	2.530	3.233	2.800	3.533	4.175	4.476	5.147
9	2.596	3.308	2.870	3.604	4.242	4.543	5.214
12	2.663	3.375	2.941	3.675	4.309	4.610	5.280
18	2.788	3.500	3.066	3.800	4.429	4.735	5.406
23	2.888	3.595	3.166	3.900	4.525	4.830	5.497
24 + under-drain	2.909	3.616	3.187	3.923	4.550	4.855	5.518

Bed Depth (inches)	Filtration Time (hours) 10.00 (9.50)	11.83 (11.33)	12.50 (12.00)
Surface	0.000	0.000	0.000
1	3.794	4.857	5.006
2	5.786	7.287	7.619
3	5.952	7.520	7.894
6	6.032	7.620	8.006
9	6.103	7.687	8.073
12	6.169	7.753	8.144
18	6.328	7.878	8.273
23	6.390	7.970	8.365
24 + under-drain	6.411	7.995	8.389

APPENDIX A (Continued)

Run No. 4 - Filter No. 2

Bed Depth (inches)	Filtration Time (hours) 0.00	0.50	1.00	1.50	2.00	3.00	4.00
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.029	0.192	0.370	0.591	0.808	1.271	1.721
2	0.071	0.225	0.391	0.633	0.879	1.433	2.017
3	0.104	0.250	0.416	0.666	0.925	1.458	2.054
6	0.196	0.334	0.495	0.741	0.992	1.533	2.121
9	0.271	0.409	0.570	0.821	1.067	1.616	2.221
12	0.346	0.475	0.637	0.891	1.133	1.687	2.288
18	0.479	0.600	0.762	1.016	1.267	1.808	2.371
23	0.575	0.700	0.854	1.108	1.358	1.908	2.508
24 + under-drain	0.596	0.721	0.870	1.129	1.383	1.933	2.537

Bed Depth (inches)	Filtration Time (hours) 5.00	6.00	*6.50 (delete)	7.00 (6.50)	7.75 (7.25)	8.17 (7.67)	8.92 (8.42)
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	2.288	2.795	2.388	3.033	3.426	3.330	3.876
2	2.692	3.420	3.155	3.904	4.476	4.518	5.164
3	2.725	3.466	3.196	3.945	4.538	4.593	5.247
6	2.792	3.475	3.275	4.025	4.613	4.672	5.331
9	2.883	3.657	3.355	4.091	4.705	4.747	5.406
12	2.938	3.662	3.421	4.158	4.759	4.813	5.472
18	3.046	3.695	3.546	4.279	4.888	4.938	5.597
23	3.162	3.875	3.642	4.366	4.984	5.034	5.693
24 + under-drain	3.196	3.966	3.667	4.391	5.017	5.059	5.718

Bed Depth (inches)	Filtration Time (hours) 10.00 (9.50)	11.83 (11.33)	12.50 (12.00)
Surface	0.000	0.000	0.000
1	4.437	5.644	5.719
2	6.011	7.561	7.852
3	6.128	7.702	7.998
6	6.216	7.811	8.085
9	6.291	7.882	8.115
12	6.357	7.944	8.256
18	6.487	8.069	8.298
23	6.582	8.165	8.473
24 + under-drain	6.607	8.186	8.498

APPENDIX A (Continued)

Run No. 4 - Filter No. 3

Bed Depth (inches)	Filtration Time (hours) 0.00	0.50	1.00	1.50	2.00	3.00	4.00
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.052	0.104	0.150	0.225	0.300	0.475	0.650
2	0.088	0.154	0.229	0.325	0.425	0.658	0.908
3	0.126	0.188	0.266	0.375	0.488	0.742	1.008
6	0.213	0.279	0.354	0.467	0.583	0.858	1.166
9	0.300	0.363	0.437	0.550	0.667	0.942	1.250
12	0.376	0.433	0.508	0.621	0.792	1.017	1.320
18	0.492	0.554	0.637	0.742	0.863	1.138	1.441
23	0.584	0.646	0.729	0.834	0.954	1.233	1.533
24 + under-drain	0.609	0.671	0.750	0.859	0.980	1.254	1.545

Bed Depth (inches)	Filtration Time (hours) 5.00	6.00	*6.50 (delete)	7.00 (6.50)	7.75 (7.25)	8.17 (7.67)	8.92 (8.42)
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.817	0.984	0.858	1.008	1.104	1.393	1.972
2	1.162	1.425	1.371	1.583	1.729	2.118	2.914
3	1.283	1.575	1.521	1.746	1.912	2.318	3.164
6	1.467	1.775	1.741	1.971	2.154	2.568	3.439
9	1.558	1.892	1.862	2.108	2.304	2.722	3.598
12	1.629	1.959	1.937	2.179	2.379	3.801	3.689
18	1.754	2.084	2.058	2.300	2.504	2.926	3.806
23	1.841	2.175	2.150	2.391	2.591	3.018	3.902
24 + under-drain	1.867	2.196	2.171	2.421	2.616	3.039	3.923

Bed Depth (inches)	Filtration Time (hours) 10.58 (10.08)	11.50 (11.00)	12.92 (12.42)
Surface	0.000	0.000	0.000
1	3.045	3.649	4.829
2	4.403	5.166	6.946
3	4.778	5.645	7.262
6	5.078	5.966	7.612
9	5.262	6.191	7.820
12	5.370	6.278	7.946
18	5.487	6.399	8.079
23	5.587	6.491	8.171
24 + under-drain	5.607	6.516	8.191

APPENDIX A (Continued)

Run No. 4 - Filter No. 4

Bed Depth (inches)	Filtration Time (hours) 0.00	0.50	1.00	1.50	2.00	3.00	4.00
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.025	0.067	0.083	0.100	0.112	0.142	0.175
2	0.059	0.108	0.141	0.167	0.191	0.250	0.292
3	0.084	0.129	0.166	0.205	0.233	0.300	0.358
6	0.154	0.208	0.241	0.284	0.312	0.392	0.471
9	0.225	0.275	0.316	0.359	0.383	0.463	0.542
12	0.300	0.350	0.391	0.434	0.454	0.534	0.609
18	0.442	0.483	0.525	0.567	0.591	0.675	0.750
23	0.534	0.579	0.625	0.663	0.683	0.767	0.842
24 + under-drain	0.567	0.600	0.641	0.688	0.704	0.788	0.867

Bed Depth (inches)	Filtration Time (hours) 5.00	6.00	*6.50 (delete)	7.00 (6.50)	7.75 (7.25)	8.17 (7.67)	8.92 (8.42)
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.196	0.225	0.146	0.158	0.200	0.372	0.693
2	0.329	0.379	0.263	0.292	0.354	0.585	0.944
3	0.396	0.459	0.325	0.362	0.433	0.676	1.044
6	0.529	0.608	0.471	0.516	0.604	0.851	1.231
9	0.600	0.692	0.571	0.616	0.712	0.968	1.352
12	0.671	0.759	0.646	0.683	0.783	1.043	1.427
18	0.804	0.896	0.779	0.825	0.921	1.180	1.560
23	0.900	0.988	0.879	0.916	1.012	1.276	1.656
24 + under-drain	0.929	1.013	0.900	0.946	1.037	1.297	1.677

APPENDIX A (Continued)

Run No. 4 - Filter No. 4 (continued)

Bed Depth (inches)	Filtration Time (hours) 10.58 (10.08)	11.50 (11.00)	12.92 (12.42)	14.50 (14.00)	15.50 (15.00)	16.50 (16.00)
Surface	0.000	0.000	0.000	0.000	0.000	0.000
1	1.182	1.590	1.933	2.737	3.259	3.737
2	1.707	2.261	2.967	3.996	4.730	5.491
3	1.899	2.411	3.146	4.217	4.984	5.782
6	2.016	2.632	3.383	4.467	5.251	6.066
9	2.207	2.778	3.538	4.629	5.417	6.237
12	2.299	2.849	3.621	4.721	5.517	6.345
18	2.391	2.990	3.758	4.854	5.651	6.478
23	2.541	3.249	3.854	4.950	5.746	6.574
24 + under-drain	2.574	3.269	3.875	4.979	5.776	6.595

*A malfunction in the manometer suppressor system resulted in a decrease in head loss at 6.50 hours during Run 4 for all filters. In order to compensate for this mishap, all data after 6.00 hours was shifted 0.50 hour toward the origin. The data for 6.50 hours was then deleted. The numbers in parenthesis are the adjusted times.

NOTE: Phosphate pump shut off after 7.25 hours of operation.

APPENDIX A (Continued)

Run No. 5 - Filter No. 1

Bed Depth (inches)	Filtration Time (hours) 0.00	1.00	2.00	3.00	4.00	5.00	6.07
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.021	0.041	0.050	0.054	0.058	0.058	0.071
2	0.046	0.091	0.125	0.150	0.175	0.191	0.221
3	0.079	0.129	0.167	0.204	0.233	0.262	0.304
6	0.158	0.212	0.275	0.337	0.387	0.433	0.492
9	0.230	0.283	0.346	0.425	0.491	0.549	0.629
12	0.291	0.345	0.408	0.491	0.683	0.633	0.721
18	0.416	0.466	0.533	0.612	0.733	0.749	0.850
23	0.508	0.562	0.629	0.704	0.775	0.841	0.946
24 + under-drain	0.533	0.583	0.649	0.729	0.795	0.870	0.967

Bed Depth (inches)	Filtration Time (hours) 7.08	8.00	9.10	10.00	11.00	12.00	13.00
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.050	0.063	0.075	0.084	0.084	0.096	0.096
2	0.200	0.238	0.271	0.292	0.330	0.363	0.387
3	0.275	0.321	0.367	0.392	0.442	0.488	0.525
6	0.475	0.538	0.604	0.655	0.717	0.788	0.842
9	0.604	0.688	0.771	0.825	0.909	0.988	1.054
12	0.700	0.796	0.888	0.950	1.042	1.134	1.204
18	0.841	0.946	1.046	1.125	1.234	1.342	1.429
23	0.933	1.042	1.146	1.234	1.346	1.454	1.554
24 + under-drain	0.958	1.067	1.171	1.255	1.371	1.479	1.575

Bed Depth (inches)	Filtration Time (hours) 14.00	15.00	16.00	17.00
Surface	0.000	0.000	0.000	0.000
1	0.117	0.125	0.137	0.158
2	0.438	0.475	0.525	0.587
3	0.583	0.634	0.695	0.779
6	0.925	0.996	1.087	1.204
9	1.150	1.238	1.341	1.475
12	1.317	1.417	1.525	1.667
18	1.550	1.667	1.800	1.958
23	1.683	1.813	1.950	2.121
24 + under-drain	1.708	1.842	1.987	2.154

APPENDIX A (Continued)

Run No. 5 - Filter No. 2

Bed Depth (inches)	Filtration Time (hours) 0.00	1.00	2.00	3.00	4.00	5.00	6.07
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.042	0.100	0.138	0.163	0.179	0.200	0.228
2	0.075	0.150	0.221	0.262	0.292	0.325	0.369
3	0.104	0.191	0.262	0.317	0.354	0.396	0.452
6	0.191	0.279	0.350	0.421	0.471	0.529	0.602
9	0.271	0.349	0.425	0.496	0.550	0.613	0.686
12	0.341	0.416	0.495	0.562	0.617	0.677	0.752
18	0.466	0.538	0.620	0.679	0.742	0.804	0.869
23	0.566	0.645	0.720	0.771	0.842	0.904	0.969
24 + under-drain	0.585	0.674	0.749	0.800	0.871	0.929	0.990

Bed Depth (inches)	Filtration Time (hours) 7.05	8.00	9.10	10.00	11.00	12.00	13.00
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.234	0.262	0.292	0.317	0.337	0.371	0.400
2	0.392	0.437	0.492	0.529	0.562	0.629	0.675
3	0.483	0.537	0.592	0.642	0.683	0.758	0.817
6	0.650	0.712	0.788	0.850	0.904	0.996	1.075
9	0.742	0.820	0.909	0.979	1.045	1.154	1.246
12	0.808	0.894	0.984	1.063	1.133	1.254	1.358
18	0.933	1.020	1.109	1.188	1.258	1.379	1.500
23	1.033	1.120	1.205	1.283	1.354	1.479	1.600
24 + under-drain	1.058	1.145	1.234	1.313	1.383	1.504	1.633

Bed Depth (inches)	Filtration Time (hours) 14.00	15.00	16.00	17.00
Surface	0.000	0.000	0.000	0.000
1	0.430	0.467	0.517	0.566
2	0.730	0.800	0.880	0.966
3	0.884	0.967	1.067	1.171
6	1.150	1.250	1.375	1.496
9	1.330	1.442	1.575	1.712
12	1.442	1.571	1.709	1.862
18	1.580	1.713	1.814	2.021
23	1.675	1.809	1.950	2.116
24 + under-drain	1.700	1.834	1.971	2.146

APPENDIX A (Continued)

Run No. 5 - Filter No. 3

Bed Depth (inches)	Filtration Time (hours) 0.00	1.00	2.00	3.00	4.00	5.00	6.07
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.041	0.151	0.200	0.214	0.259	0.283	0.333
2	0.083	0.226	0.309	0.346	0.425	0.479	0.550
3	0.116	0.251	0.351	0.408	0.504	0.576	0.675
6	0.187	0.321	0.426	0.500	0.617	0.716	0.850
9	0.262	0.392	0.496	0.566	0.688	0.787	0.933
12	0.328	0.459	0.563	0.638	0.754	0.854	1.000
18	0.466	0.584	0.692	0.758	0.879	0.979	1.121
23	0.557	0.688	0.792	0.858	0.979	1.079	1.225
24 + under-drain	0.583	0.709	0.817	0.879	1.000	1.104	1.250

Bed Depth (inches)	Filtration Time (hours) 7.02	8.00	9.10	*10.00 (delete)	11.00 (9.50)	12.00 (10.50)	13.00 (11.50)
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.333	0.384	0.446	0.458	0.446	0.516	0.554
2	0.567	0.659	0.750	0.775	0.763	0.887	0.946
3	0.691	0.808	0.933	0.963	0.946	1.058	1.188
6	0.883	1.017	1.183	1.217	1.204	1.391	1.492
9	0.983	1.133	1.317	1.363	1.354	1.566	1.683
12	1.050	1.208	1.400	1.454	1.454	1.675	1.808
18	1.175	1.329	1.533	1.583	1.584	1.816	1.950
23	1.275	1.433	1.633	1.692	1.688	1.921	2.054
24 + under-drain	1.299	1.454	1.654	1.713	1.717	1.941	2.075

Bed Depth (inches)	Filtration Time (hours) 14.00 (12.50)	15.00 (13.50)	16.00 (14.50)	17.00 (15.50)
Surface	0.000	0.000	0.000	0.000
1	0.592	0.662	0.684	0.750
2	1.017	1.125	1.163	1.258
3	1.288	1.416	1.475	1.621
6	1.617	1.775	1.838	2.008
9	1.825	2.008	2.075	2.262
12	1.958	2.154	2.230	2.437
18	2.121	2.329	2.413	2.637
23	2.225	2.441	2.513	2.746
24 + under-drain	2.246	2.462	2.542	2.766

APPENDIX A (Continued)

Run No. 5 - Filter No. 4

Bed Depth (inches)	Filtration Time (hours) 0.00	1.00	2.00	3.00	4.00	5.00	6.05
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.041	0.046	0.058	0.074	0.076	0.083	0.087
2	0.077	0.088	0.108	0.122	0.134	0.150	0.158
3	0.102	0.121	0.137	0.158	0.175	0.183	0.200
6	0.194	0.213	0.241	0.266	0.288	0.308	0.333
9	0.236	0.283	0.308	0.341	0.375	0.396	0.425
12	0.306	0.350	0.379	0.420	0.459	0.479	0.516
18	0.423	0.483	0.508	0.553	0.592	0.625	0.662
23	0.519	0.579	0.608	0.649	0.692	0.720	0.766
24 + under-drain	0.553	0.613	0.637	0.716	0.721	0.750	0.791

Bed Depth (inches)	Filtration Time (hours) 7.00	8.00	9.10	10.00	11.00	12.00	13.00
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.100	0.104	0.108	0.116	0.121	0.133	0.146
2	0.174	0.187	0.196	0.212	0.225	0.241	0.263
3	0.224	0.237	0.254	0.275	0.292	0.312	0.333
6	0.341	0.387	0.408	0.433	0.459	0.491	0.525
9	0.470	0.495	0.529	0.562	0.592	0.629	0.667
12	0.558	0.587	0.629	0.666	0.700	0.750	0.788
18	0.716	0.749	0.792	0.841	0.880	0.933	0.983
23	0.816	0.858	0.908	0.958	1.000	1.058	1.108
24 + under-drain	0.849	0.887	0.938	0.987	1.030	1.087	1.142

APPENDIX A (Continued)

Run No. 5 - Filter No 4. (continued)

Bed Depth (inches)	Filtration Time (hours) 14.00	15.00	16.00	17.00
Surface	0.000	0.000	0.000	0.000
1	0.158	0.171	0.183	0.191
2	0.283	0.304	0.325	0.345
3	0.358	0.387	0.408	0.437
6	0.558	0.596	0.621	0.662
9	0.700	0.746	0.783	0.829
12	0.825	0.875	0.917	0.962
18	1.033	1.087	1.133	1.195
23	1.158	1.221	1.275	1.337
24 + under-drain	1.196	1.258	1.308	1.370

*A malfunction in the manometer suppressor system resulted in disturbance of the filter bed in Filter Number 3 during Run 5. In order to compensate for this mishap, all data after 9.10 hours was shifted 1.5 hours toward the origin. The data for 10.00 hours was deleted. The numbers in parenthesis are the adjusted times.

APPENDIX A (Continued)

Run No. 6 - Filter No. 1

Bed Depth (inches)	Filtration Time (hours) 0.00	1.00	2.00	3.00	4.00	5.00	6.02
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.012	0.450	0.996	1.608	2.100	2.571	3.142
2	0.041	0.533	1.278	2.200	3.096	3.925	4.834
3	0.066	0.558	1.305	2.216	3.121	3.975	4.946
6	0.137	0.629	1.380	2.291	3.192	4.050	5.021
9	0.208	0.700	1.442	2.358	3.267	4.125	5.088
12	0.274	0.775	1.530	2.433	3.338	4.192	5.159
18	0.408	0.904	1.659	2.566	3.471	4.321	5.288
23	0.500	0.996	1.755	2.654	3.563	4.421	5.379
24 + under-drain	0.516	1.016	1.775	2.675	3.579	4.442	5.400

Bed Depth (inches)	Filtration Time (hours) 8.62	9.52	10.50
Surface	0.000	0.000	0.000
1	3.368	3.601	4.163
2	6.772	7.564	8.830
3	7.114	8.031	9.321
6	7.218	8.168	9.488
9	7.293	8.243	9.555
12	7.359	8.314	9.621
18	7.493	8.443	9.751
23	7.593	8.535	9.855
24 + under-drain	7.614	8.556	9.871

Run No. 6 - Filter No. 2

Bed Depth (inches)	Filtration Time (hours) 0.00	1.00	2.00	3.00	4.00	5.58	7.92
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.117	0.870	1.746	1.925	3.737	4.837	7.687
2	0.150	1.008	2.121	3.525	4.916	6.941	10.854
3	0.179	1.037	2.146	3.554	4.966	7.041	11.075
6	0.263	1.120	2.229	3.637	5.049	7.128	11.216
9	0.338	1.195	2.300	3.712	5.124	7.203	11.291
12	0.408	1.262	2.367	3.779	5.187	7.270	11.358
18	0.525	1.387	2.492	3.896	5.308	7.387	11.479
23	0.617	1.474	2.584	3.987	5.395	7.474	11.571
24 + under-drain	0.637	1.495	2.604	4.008	5.416	7.495	11.591

APPENDIX A (Continued)

Run No. 6 - Filter No. 3

Bed Depth (inches)	Filtration Time (hours) 0.00	1.00	2.00	3.00	4.00	5.00	6.00
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.033	0.209	0.396	0.604	0.825	1.112	1.762
2	0.067	0.317	0.583	0.908	1.267	1.708	2.546
3	0.096	0.350	0.663	1.037	1.458	1.995	2.929
6	0.171	0.425	0.746	1.154	1.621	2.212	3.179
9	0.246	0.500	0.821	1.229	1.696	2.291	3.275
12	0.317	0.567	0.896	1.300	1.767	2.362	3.346
18	0.446	0.696	1.013	1.425	1.892	2.487	3.471
23	0.538	0.792	1.109	1.520	1.987	2.583	3.566
24 + under-drain	0.559	0.813	1.133	1.541	2.008	2.608	3.591

Bed Depth (inches)	Filtration Time (hours) 7.50	8.97	10.00
Surface	0.000	0.000	0.000
1	2.654	3.479	3.826
2	4.046	5.404	6.476
3	4.579	5.950	6.989
6	4.866	6.291	7.368
9	5.016	6.471	7.568
12	5.087	6.554	7.676
18	5.216	6.679	7.801
23	5.312	6.779	7.893
24 + under-drain	5.333	6.800	7.918

APPENDIX A (Concluded)

Run No. 6 - Filter No. 4

Bed Depth (inches)	Filtration Time (hours) 0.00	1.00	2.00	3.00	4.00	5.00	5.98
Surface	0.000	0.000	0.000	0.000	0.000	0.000	0.000
1	0.034	0.113	0.133	0.159	0.175	0.200	0.233
2	0.067	0.188	0.212	0.242	0.192	0.308	0.350
3	0.092	0.217	0.262	0.300	0.333	0.375	0.421
6	0.167	0.300	0.379	0.438	0.479	0.533	0.583
9	0.246	0.375	0.458	0.542	0.600	0.662	0.721
12	0.317	0.450	0.533	0.617	0.700	0.771	0.837
18	0.450	0.584	0.666	0.755	0.833	0.921	1.016
23	0.550	0.679	0.758	0.850	0.929	1.017	1.108
24 + under-drain	0.571	0.704	0.783	0.875	0.954	1.037	1.133

Bed Depth (inches)	Filtration Time (hours) 7.50	9.13	10.08
Surface	0.000	0.000	0.000
1	0.242	0.215	0.300
2	0.379	0.419	0.521
3	0.454	0.432	0.625
6	0.592	0.711	0.830
9	0.775	0.871	0.992
12	0.900	1.007	1.142
18	1.113	1.240	1.380
23	1.217	1.369	1.517
24 + under-drain	1.242	1.393	1.550

APPENDIX B

Penetration Data

Run Number 3
Visual Penetration (Inches)

Filtration Time (hours)	Filter 1	Filter 2	Filter 3	Filter 4
0.00	0.0	0.0	0.0	0.0
5.35	2.3	3.2	1.5	2.1
7.45	2.8	4.4	2.0	2.9
9.50	3.2	5.5	2.5	3.5
10.08	3.4	5.9	2.6	3.8
11.45	3.8	6.5	3.0	4.0

NOTE: Phosphate pump operated only from 2:00 to 3:00 hours.

Run Number 4
Visual Penetration (Inches)

Filtration Time (hours)	Filter 1	Filter 2	Filter 3	Filter 4
0.00	0.0	0.0	0.0	0.0
0.25	0.4	0.2	0.5	0.8
0.58	0.5	0.3	1.2	1.2
0.80	0.6	0.4	1.8	2.0
1.13	0.7	0.8	2.9	2.6
2.15	0.8	0.9	4.8	4.1
3.05	1.0	1.4	5.9	4.9
4.20	1.4	1.9	6.8	6.2
6.25	2.0	2.5	9.0	8.2
7.40	-	-	-	8.8
7.83	-	-	-	9.0
8.25	-	-	-	9.0
8.33	2.5	2.9	11.0	-
9.25	-	-	-	9.1
10.12	3.2	3.4	12.5	9.1
11.92	3.5	3.6	12.8	10.0
13.33	4.0	4.0	14.2	10.7
14.50	-	-	-	11.1
15.58	-	-	-	11.9
16.67	-	-	-	12.2

NOTE: Phosphate pump shut off after 7.25 hours of operation.

APPENDIX B (Concluded)

Run Number 5
Visual Penetration
(Inches)

Filtration Time (hours)	Filter 1	Filter 2	Filter 3	Filter 4
0.00	0.0	0.0	0.0	0.0
0.50	1.9	1.5	1.1	2.9
1.00	3.2	2.2	1.9	7.0
1.50	4.0	3.0	2.5	7.5
2.17	6.2	3.5	3.1	8.4
3.17	9.1	5.0	4.5	11.0
4.07	10.8	5.9	5.8	15.2
5.05	13.2	7.0	7.0	17.5
6.13	14.8	8.4	8.6	19.5
7.08	17.6	8.6	9.6	21.1
8.05	18.2	9.6	10.0	22.5
9.20	20.2	11.2	12.1	24.0
10.07	20.4	11.8	12.8	-
10.92	21.1	12.0	14.1	-
12.08	22.2	13.1	16.0	-
13.07	23.5	14.4	16.9	-
14.08	24.0	14.4	17.0	-
15.08	-	14.7	17.2	-
16.08	-	14.9	17.9	-
17.07	-	15.5	18.2	-

Run Number 6
Visual Penetration
(Inches)

Filtration Time (hours)	Filter 1	Filter 2	Filter 3	Filter 4
0.00	0.0	0.0	0.0	0.0
0.08	0.2	0.2	0.4	0.8
0.58	0.6	0.8	2.0	2.5
1.17	0.8	1.1	2.8	5.0
1.50	0.9	1.2	2.9	5.4
2.05	1.8	2.0	3.5	7.0
2.92	1.8	2.1	4.4	9.2
3.97	2.8	2.9	5.5	11.8
4.97	3.0	3.2	6.8	14.3
5.92	3.1	4.0	7.1	16.2
8.00	3.8	5.5	9.2	20.8
9.50	4.2	-	10.2	24.0
10.55	5.2	-	12.0	-

APPENDIX C

Iron Data

Run Number 4
Influent Iron Concentration
(mg/l)

Filtration Time (hours)	Filter 1	Filter 2	Filter 3	Filter 4
0.00	-	1.66	2.68	2.60
1.08	2.30	2.74	6.86	4.34
2.00	2.36	2.58	6.86	4.24
4.08	2.72	2.56	6.84	3.96
6.17	3.38	3.12	8.12	4.00
8.42	4.08	2.96	7.46	3.46
10.13	3.86	-	6.98	4.26
12.00	3.50	2.70	6.28	4.30
13.00	-	-	-	4.76
13.05	-	2.52	-	-
13.10	-	-	6.56	-
13.15	3.06	-	-	-
14.50	-	-	-	4.44
15.35	-	-	-	4.70
16.33	-	-	-	4.94
Average	3.16	2.60	6.52	4.17

Run Number 5
Influent Iron Concentration
(mg/l)

Filtration Time (hours)	Filter 1	Filter 2	Filter 3	Filter 4
0.92	5.14	2.56	3.84	3.36
2.10	6.24	2.60	3.80	3.40
3.20	6.54	2.72	3.86	3.72
4.10	6.34	2.72	-	4.20
5.15	6.00	2.84	4.60	3.72
6.17	5.74	3.14	4.54	3.84
7.10	5.34	3.06	4.26	3.84
8.05	6.46	3.48	4.22	4.08
9.25	5.52	2.84	4.26	4.12
10.13	5.52	2.92	4.20	4.10
11.08	5.50	3.00	4.20	4.06

(continued)

APPENDIX C (Concluded)

Run Number 5 (continued)

Filtration Time (hours)	Filter 1	Filter 2	Filter 3	Filter 4
12.10	5.72	3.10	4.26	4.24
1.08	5.56	3.20	4.20	4.20
2.13	5.56	3.14	4.32	4.20
3.10	5.56	3.24	4.32	4.32
4.20	5.42	3.44	4.26	4.20
5.12	5.20	3.24	4.04	4.08
Average	5.73	3.01	4.20	3.98

Run Number 6
Influent Iron Concentration (mg/l)

Filtration Time (hours)	Filter 1	Filter 2	Filter 3	Filter 4
1.07	2.96	3.48	4.44	4.20
2.00	3.54	3.58	4.26	4.32
3.08	4.06	4.44	4.70	5.12
4.10	3.48	3.74	4.56	4.64
5.08	7.56	3.44	4.56	4.52
6.03	3.86	3.68	4.90	5.56
7.00	5.12	4.84	7.08	4.84
8.00	4.56	5.04	6.98	1.08
9.00	5.66	-	7.44	4.84
10.22	3.76	-	8.84	5.74
Average	4.46	4.02	5.78	4.49

APPENDIX D

Total Iron Applied Versus Head Loss

Run Number 4

Filtration Time (hours)	Filter 1 Total Iron Appl'd (mgFe)	Filter 1 Total Head Loss (ft)	Filter 2 Total Iron Appl'd (mgFe)	Filter 2 Total Head Loss (ft)	Filter 3 Total Iron Appl'd (mgFe)	Filter 3 Total Head Loss (ft)	Filter 4 Total Iron Appl'd (mgFe)	Filter 4 Total Head Loss (ft)
0.00	0.00	0.550	0.00	0.596	0.00	0.609	0.00	0.567
0.50	35.24	0.691	29.07	0.721	72.72	0.671	46.51	0.600
1.00	70.48	0.841	58.14	0.870	145.44	0.750	93.01	0.641
1.50	105.72	1.083	87.21	1.129	218.16	0.859	139.51	0.688
2.00	140.96	1.262	116.28	1.383	290.87	0.980	186.02	0.704
3.00	211.44	1.750	174.42	1.933	436.31	1.254	279.03	0.788
4.00	-	-	232.56	2.537	581.75	1.545	372.04	0.867
5.00	352.40	2.909	290.70	3.196	727.18	1.867	465.05	0.929
6.00	422.88	3.616	348.84	3.966	872.62	2.196	558.06	1.013
6.50	458.12	3.925	377.91	4.391	945.34	2.421	-	-
7.25	-	-	-	-	-	-	614.32	1.037
7.67	540.37	4.855	445.76	5.059	1115.07	3.039	713.11	1.297
8.42	593.23	5.518	489.36	5.718	1224.14	3.923	782.87	1.667
9.50	669.56	6.411	552.33	6.607	-	-	-	-
10.08	-	-	-	-	1466.44	5.607	957.82	2.574
11.00	-	-	-	-	1599.81	6.516	1023.11	3.269
11.33	798.75	7.995	658.90	8.186	-	-	-	-
12.00	845.76	8.389	697.68	8.498	-	-	-	-
12.42	-	-	-	-	1805.89	8.191	1154.91	3.875
14.00	-	-	-	-	-	-	1302.14	4.979
15.00	-	-	-	-	-	-	1395.15	5.776
16.00	-	-	-	-	-	-	1488.18	6.595

See NOTE for Run 4 in Appendix A.

APPENDIX D (Continued)

Run Number 5

Filtration Time (hours)	Filter 1 Total Iron Appl'd (mgFe)	Filter 1 Total Head Loss (ft)	Filter 2 Total Iron Appl'd (mgFe)	Filter 2 Total Head Loss (ft)	Filter 3 Total Iron Appl'd (mgFe)	Filter 3 Total Head Loss (ft)	Filter 4 Total Iron Appl'd (mgFe)	Filter 4 Total Head Loss (ft)
0.00	0.00	0.533	0.00	0.585	0.00	0.583	0.00	0.553
1.00	127.80	0.583	67.27	0.674	93.72	0.709	88.86	0.613
2.00	255.60	0.649	134.55	0.749	187.44	0.817	177.71	0.637
3.00	383.40	0.729	201.82	0.800	281.17	0.879	266.57	0.716
4.00	511.20	0.795	269.09	0.871	374.89	1.000	355.42	0.721
5.00	639.00	0.870	336.37	0.929	468.61	1.104	444.28	0.750
6.05	-	-	-	-	-	-	537.58	0.791
6.07	-	-	408.15	0.990	568.61	1.250	-	-
7.00	-	-	-	-	-	-	621.99	0.849
7.02	-	-	-	-	657.65	1.299	-	-
7.05	-	-	474.27	1.058	-	-	-	-
7.08	905.21	0.958	-	-	-	-	-	-
8.00	1022.40	1.067	538.18	1.145	749.78	1.454	710.85	0.887
9.10	1162.98	1.171	612.18	1.234	852.87	1.654	808.59	0.938
9.50	-	-	-	-	890.36	1.717	-	-
10.00	1278.00	1.255	672.73	1.313	-	-	888.56	0.987
10.50	-	-	-	-	984.08	1.941	-	-
11.00	1405.84	1.371	740.00	1.383	-	-	977.42	1.030
11.50	-	-	-	-	1077.80	2.075	-	-
12.00	1533.65	1.479	807.28	1.504	-	-	1066.27	1.087
12.50	-	-	-	-	1171.53	2.246	-	-
13.00	1661.45	1.575	874.55	1.633	-	-	1155.13	1.142
13.50	-	-	-	-	1265.25	2.462	-	-
14.00	1789.26	1.708	941.82	1.700	-	-	1243.98	1.196
14.50	-	-	-	-	1358.97	2.542	-	-
15.00	1917.06	1.842	1009.09	1.834	-	-	1332.84	1.258
15.50	-	-	-	-	1452.69	2.766	-	-
16.00	2044.86	1.987	1076.37	1.971	-	-	1421.70	1.308
17.00	2172.67	2.154	1143.64	2.146	-	-	1510.55	1.370

See NOTE for Run 5, Filter 3 in Appendix A

APPENDIX D (Concluded)

Run Number 6

Filtration Time (hours)	Filter 1 Total Iron Appl'd (mgFe)	Filter 1 Total Head Loss (ft)	Filter 2 Total Iron Appl'd (mgFe)	Filter 2 Total Head Loss (ft)	Filter 3 Total Iron Appl'd (mgFe)	Filter 3 Total Head Loss (ft)	Filter 4 Total Iron Appl'd (mgFe)	Filter 4 Total Head Loss (ft)
0.00	0.00	0.516	0.00	0.637	0.00	0.559	0.00	0.571
1.00	99.46	1.016	89.84	1.495	128.92	0.813	100.13	0.704
2.00	198.92	1.775	179.68	2.604	257.84	1.133	200.26	0.783
3.00	298.38	2.675	269.52	4.008	386.76	1.541	300.39	0.875
4.00	397.84	3.579	359.36	5.416	515.68	2.008	400.52	0.954
5.00	497.30	4.442	-	-	644.60	2.608	500.65	1.037
5.58	-	-	501.58	7.495	-	-	-	-
5.98	-	-	-	-	-	-	599.08	1.133
6.00	-	-	-	-	773.52	3.591	-	-
6.02	598.45	5.400	-	-	-	-	-	-
7.50	-	-	-	-	966.90	5.333	750.97	1.242
7.92	-	-	711.26	11.591	-	-	-	-
8.62	857.05	7.614	-	-	-	-	-	-
8.97	-	-	-	-	1156.03	6.800	-	-
9.13	-	-	-	-	-	-	914.49	1.393
9.52	946.36	8.556	-	-	-	-	-	-
10.00	-	-	-	-	1289.20	7.918	-	-
10.08	-	-	-	-	-	-	1009.61	1.550
10.50	1044.33	9.871	-	-	-	-	-	-

APPENDIX E

Total Iron Applied Versus Penetration

Run Number 4

Filtration Time (hours)	Filter 1 Total Iron Appl'd (mgFe)	Filter 1 Penetration (in)	Filter 2 Total Iron Appl'd (mgFe)	Filter 2 Penetration (in)	Filter 3 Total Iron Appl'd (mgFe)	Filter 3 Penetration (in)	Filter 4 Total Iron Appl'd (mgFe)	Filter 4 Penetration (in)
0.00	0.00	0.0	0.00	0.0	0.00	0.0	0.00	0.0
0.25	17.62	0.4	14.53	0.2	36.36	0.5	23.25	0.8
0.58	41.10	0.5	33.90	0.3	84.79	1.2	54.22	1.2
0.80	56.39	0.6	46.51	0.4	116.35	1.8	74.41	2.0
1.13	79.87	0.7	65.87	0.8	164.78	2.9	105.38	2.6
2.15	151.55	0.8	125.00	0.9	312.69	4.8	199.97	4.1
3.05	214.99	1.0	177.32	1.4	443.58	5.9	283.68	4.9
4.20	296.06	1.4	244.19	1.9	610.84	6.8	390.64	6.2
6.25	440.56	2.0	363.37	2.5	908.98	9.0	581.31	8.2
7.40	-	-	-	-	-	-	688.27	8.8
7.83	-	-	-	-	-	-	728.55	9.0
8.25	-	-	-	-	-	-	767.33	9.0
8.33	587.39	2.5	484.48	2.9	1211.93	11.0	-	-
9.25	-	-	-	-	-	-	860.34	9.1
10.12	713.08	3.2	588.14	3.4	1471.24	12.5	940.89	9.1
11.92	840.03	3.5	692.85	3.6	1733.17	12.8	1108.40	10.0
13.33	939.84	4.0	775.18	4.0	1939.11	14.2	1240.10	10.7
14.50	-	-	-	-	-	-	1348.65	11.1
15.58	-	-	-	-	-	-	1449.37	11.9
16.67	-	-	-	-	-	-	1530.20	12.2

Phosphate pump shut off after 7.25 hours of operation.

APPENDIX E (Continued)

Run Number 5

Filtration Time (hours)	Filter 1 Total Iron Appl'd (mgFe)	Filter 1 Penetration (in)	Filter 2 Total Iron Appl'd (mgFe)	Filter 2 Penetration (in)	Filter 3 Total Iron Appl'd (mgFe)	Filter 3 Penetration (in)	Filter 4 Total Iron Appl'd (mgFe)	Filter 4 Penetration (in)
0.00	0.00	0.0	0.00	0.0	0.00	0.0	0.00	0.0
0.50	63.90	1.9	33.64	1.5	46.86	1.1	44.43	2.9
1.00	127.80	3.2	67.27	2.2	93.72	1.9	88.86	7.0
1.50	191.71	4.0	100.91	3.0	140.58	2.5	133.28	7.5
2.17	276.95	6.2	145.78	3.5	203.10	3.1	192.55	8.4
3.17	404.76	9.1	213.05	5.0	296.82	4.5	281.41	11.0
4.07	519.78	10.8	273.60	5.9	381.17	5.8	361.38	15.2
5.05	645.41	13.2	339.72	7.0	473.30	7.0	448.72	17.5
6.13	783.82	14.8	412.58	8.4	574.80	8.6	544.95	19.5
7.08	905.24	17.6	476.49	8.6	663.83	9.6	629.37	21.1
8.05	1028.82	18.2	541.54	9.6	754.46	10.0	715.29	22.5
9.20	1175.80	20.2	618.90	11.2	862.24	12.1	817.48	24.0
10.07	1286.60	20.4	677.23	11.8	943.50	12.8	-	-
10.92	1395.24	21.1	734.41	12.0	1023.16	14.1	-	-
12.08	1544.26	22.2	812.85	13.1	1132.44	16.0	-	-
13.07	1670.01	23.5	879.04	14.4	1224.67	16.9	-	-
14.08	1799.86	24.0	947.39	14.4	1319.89	17.0	-	-
15.08	-	-	1014.66	14.7	1413.61	17.2	-	-
16.08	-	-	1081.94	14.9	1507.33	17.9	-	-
17.07	-	-	1148.13	15.5	1599.55	18.2	-	-

APPENDIX E (Concluded)

Run Number 6

Filtration Time (hours)	Filter 1 Total Iron Appl'd (mgFe)	Filter 1 Penetration (in)	Filter 2 Total Iron Appl'd (mgFe)	Filter 2 Penetration (in)	Filter 3 Total Iron Appl'd (mgFe)	Filter 3 Penetration (in)	Filter 4 Total Iron Appl'd (mgFe)	Filter 4 Penetration (in)
0.00	0.00	0.0	0.00	0.0	0.00	0.0	0.00	0.0
0.08	8.26	0.2	7.46	0.2	10.70	0.4	8.31	0.8
0.58	57.99	0.6	52.38	0.8	75.16	2.0	58.38	2.5
1.17	116.07	0.8	104.84	1.1	150.45	2.8	116.85	5.0
1.50	149.19	0.9	134.76	1.2	193.38	2.9	150.19	5.4
2.05	203.89	1.8	184.17	2.0	264.29	3.5	205.27	7.0
2.92	290.03	1.8	261.97	2.1	375.93	4.4	291.98	9.2
3.97	394.56	2.8	356.40	2.9	511.43	5.5	397.22	11.8
4.97	494.02	3.0	446.24	3.2	640.35	6.8	497.35	14.3
5.92	588.41	3.1	531.49	4.0	762.69	7.1	592.37	16.2
8.00	795.68	3.8	718.72	5.5	1031.36	9.2	801.04	20.8
9.50	941.87	4.2	-	-	1224.74	10.2	951.24	24.0
10.55	1049.30	5.2	-	-	1360.10	12.0	-	-

APPENDIX F

Head Loss per Inch of Filter Depth Versus Time

Run No. 4 - Filter No. 4

Bed Depth Interval (inches)	Filtration Time (hours) 0.00	2.00	4.00	6.00	8.42	12.42
0-1	0.025	0.112	0.175	0.225	0.693	1.933
1-2	0.034	0.079	0.117	0.154	0.251	1.034
2-3	0.025	0.042	0.066	0.080	0.100	0.179
3-6	0.023	0.023	0.028	0.049	0.062	0.079
6-9	0.024	0.024	0.024	0.026	0.040	0.052
9-12	0.025	0.024	0.022	0.022	0.025	0.028
12-18	0.024	0.023	0.024	0.023	0.022	0.023
18-23	0.018	0.018	0.018	0.018	0.019	0.019
23-24+ Underdrain	0.033	0.021	0.025	0.025	0.021	0.021
6-24+ Underdrain	0.025	0.022	0.023	0.023	0.025	0.029

Run No 4. - Filter No. 4 (continued)

Bed Depth Interval (inches)	Filtration Time (hours) 14.00	16.00
0-1	2.737	3.737
1-2	1.259	1.754
2-3	0.221	0.291
3-6	0.083	0.095
6-9	0.054	0.057
9-12	0.031	0.039
12-18	0.022	0.024
18-23	0.019	0.019
23-24+ Underdrain	0.029	0.021
6-24+ Underdrain	0.031	0.033

APPENDIX F (Continued)

Run No. 6 - Filter No. 2

Bed Depth Interval (inches)	Filtration Time (hours) 0.00	1.00	2.00	4.00	5.58	7.92
0-1	0.117	0.870	1.746	3.737	4.837	7.687
1-2	0.033	0.138	0.375	1.179	2.104	3.167
2-3	0.029	0.025	0.025	0.050	0.100	0.221
3-6	0.028	0.028	0.028	0.028	0.029	0.047
6-9	0.025	0.025	0.024	0.025	0.025	0.025
9-12	0.023	0.022	0.022	0.021	0.022	0.022
12-18	0.019	0.021	0.021	0.020	0.019	0.020
18-23	0.018	0.017	0.018	0.018	0.018	0.018
23-24 + Underdrain	0.020	0.021	0.020	0.021	0.021	0.020
3-24 + Underdrain	0.022	0.022	0.022	0.022	0.022	0.025

Run No. 5 - Filter No. 4

Bed Depth Interval (inches)	Filtration Time (hours) 0.00	2.00	4.00	6.05	8.00	10.00
0-1	0.041	0.058	0.076	0.087	0.104	0.116
1-2	0.036	0.050	0.058	0.071	0.083	0.096
2-3	0.025	0.029	0.041	0.042	0.050	0.063
3-6	0.031	0.034	0.051	0.044	0.050	0.053
6-9	0.014	0.022	0.029	0.031	0.036	0.043
9-12	0.023	0.023	0.028	0.030	0.031	0.035
12-18	0.019	0.021	0.022	0.024	0.027	0.029
18-23	0.019	0.020	0.020	0.021	0.022	0.023
23-24 + Underdrain	0.034	0.029	0.029	0.025	0.029	0.029
9-24 + Underdrain	0.024	0.023	0.025	0.025	0.027	0.029

Run No. 5 - Filter No. 4 (continued)

Bed Depth Interval (inches)	Filtration Time (hours) 12.00	14.00	17.00
0-1	0.133	0.158	0.191
1-2	0.108	0.125	0.154
2-3	0.071	0.075	0.092
3-6	0.063	0.067	0.075
6-9	0.046	0.047	0.056
9-12	0.040	0.042	0.044
12-18	0.030	0.035	0.039
18-23	0.025	0.025	0.028
23-24 + Underdrain	0.029	0.038	0.033
9-24 + Underdrain	0.031	0.035	0.036

APPENDIX F (Concluded)

Run No. 5 - Filter No. 1

Bed Depth Interval (inches)	Filtration Time (hours) 0.00	2.00	4.00	6.07	8.00	10.00
0-1	0.021	0.050	0.058	0.071	0.063	0.084
1-2	0.025	0.075	0.117	0.150	0.175	0.208
2-3	0.033	0.042	0.058	0.083	0.083	0.100
3-6	0.026	0.036	0.051	0.063	0.072	0.088
6-9	0.024	0.023	0.035	0.046	0.050	0.057
9-12	0.020	0.021	0.070	0.031	0.036	0.042
12-18	0.021	0.021	0.064	0.022	0.025	0.029
18-23	0.018	0.019	0.018	0.019	0.019	0.022
23-24+ Underdrain	0.025	0.020	0.020	0.021	0.025	0.021
9-24+ Underdrain	0.021	0.020	0.043	0.023	0.026	0.028

Run No. 5 - Filter No. 1 (continued)

Bed Depth Interval (inches)	Filtration Time (hours) 12.00	14.00	17.00
0-1	0.096	0.117	0.158
1-2	0.267	0.321	0.429
2-3	0.125	0.145	0.192
3-6	0.100	0.114	0.142
6-9	0.067	0.078	0.090
9-12	0.049	0.056	0.064
12-18	0.035	0.039	0.048
18-23	0.022	0.027	0.033
23-24+ Underdrain	0.025	0.025	0.033
9-24+ Underdrain	0.033	0.037	0.044

APPENDIX G

The Effect of pH on the Electrophoretic Mobility of Silica

pH	Mobility μ/sec/volt/cm
1.27	$\pm$ 0
2.03	- 0.75
2.98	- 1.31
3.91	- 1.82
4.27	- 1.67
6.18	- 2.09
6.90	- 2.56
7.02	- 2.56
9.50	- 3.15
9.62	- 3.38
10.40	- 3.42
11.92	- 3.99

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